The Continued Study of Spouted Bed Gasification of Western Canadian Coal

Final Report
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by

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ABSTRACT

This report describes work carried out in Fiscal Year 1982-83. Experimental results on the spouted bed gasifier are reported for one new coal - Balmer, and a high-ash coal waste product Balmer Rejects. The gasifier was modified substantially to recycle char continuously from the first solid/gas separation cyclone. The recycling of char either to the coal feed line or to the annulus of the spouted bed is essentially trouble-free and is done routinely. Effects of char recycle up to 16% of the MAF coal feed have been established. The gasifier is now equipped to run on oxygen/steam mixtures as well as air/steam mixtures and the full range of oxygen-enriched air. Results are presented for a range of temperatures, O_2/coal and O_2/steam ratios. Performance approaches that of commercially developed units, yielding gases of about 28% CO, 36% H_2 and 3% CH_4 with heating values of about 9 MJ/m^3. Data for total of twenty-four gasifier runs are reported.

1. INTRODUCTION

This report describes work done during 1982-83 on DSS Contract Serial No. OSU82-00138. The primary objective of this study is to develop the spouted bed process for gasifying Western Canadian coals. Previous work on this project has been reported (1,2,3). The specific objectives for work during 1982-83 were:

1. Extend range of coals gasified to provide broader data bank on behaviour of Western Canadian coals.

2. Operate spouted bed on oxygen rather than air to produce a synthesis gas instead of a low calorific value gas.

3. Complete the modelling study of the spouted bed gasifier, and with it the experimental measurements of in-bed composition profiles.

4. Study the feasibility of sulphur removal by injection of limestone or dolomite into the gasifier.

5. Provide backup data and information as required for the design study of the 15 tonne/day gasifier.

The breakdown of objectives 1 to 4 into a series of Major Tasks is given in the work Statement, Appendix A. This report describes the progress towards achieving the above objectives in Fiscal Year 1982-83.

2. PILOT PLANT MODIFICATIONS (Major Tasks 2 and 3)

A number of modifications to the existing pilot plant were made to complete the Major Tasks of Appendix A.

2.1 Steam System

The existing pilot plant is shown in Figures 1 and 2. The 1982-83 contract proposed to convert the unit from a low calorific value gasifier to a medium calorific value gasifier. Thus the air was to be replaced by pure oxygen and steam as the spouting fluid which would produce a nitrogen-free synthesis gas. This necessitated replacing the existing ½" steam pipe line with a 1" pipe line and buying a new steam rotameter (flowrate 1-4.13 kg/min at 35 psig) to accomodate the higher flowrate). The system (Fig. 3) consists of two solenoid valves and two steam rotameters connected in parallel. This arrangement allows either one or both branches to be used. The two solenoid valves are connected to the safety panel and are shut off automatically when there is a high pressure build up in the reactor.

2.2 Oxygen System

A new oxygen line (Fig. 4) is required to pipe the oxygen to the reactor for the synthesis gas experiments. Up to three PGS45 liquid oxygen cylinders,

each with a capacity of 127.4 m^3 gas at 1 atmosphere pressure and 20°C, can be connected in parallel. Approximately 0.34 m^3/min of oxygen is used during the experiment; a maximum of 18.7 hr. of run time is possible with three fresh liquid oxygen cylinders. But due to leakage through the cylinder pressure relief valve the actual time is much less. The cylinders are locked inside a cage which is located outside the building for easy access. The oxygen is piped through degreased ½" type L copper tube to laboratory. Depending on the experimental conditions the oxygen can be introduced at the propane burner gas inlet or at the coal feed line to transport the coal to the reactor. The solenoid is also connected to the safety panel and is shut off automatically when there is a high temperature or a high pressure detected in the reactor.

2.3 Incinerator Modification

In the original design of the incinerator, a natural gas pilot flame was thought to be sufficient to ignite the off-gas. Combustion of the gasifier product could not be maintained. This was attributed to the moisture content in the gas being too high which apparently quenched the pilot flame. Discussions and tests with the manufacturers of the incinerator failed to achieve a solution to the problem. A propane burner was therefore installed in place of the manhole cover at the stack side of the incinerator chamber (Fig. 5). This burner is self-aspirating and so no air lines was required. Propane is supplied to the burner by tapping from the existing propane line. The burner consumes about 9 kg of propane per hour. The H_2 and CO levels are reduced to less than 0.2 vol.% and the obnoxious odour is also reduced significantly.

2.4 Dust Recycling System

As reported previously (2) recycling the large amount of carryover dust would improve the carbon conversion in the gasification process. A dust recycling system was devised to feed the dust collected by the first cyclone back to the reactor. The heart of the system is the rotary valve (Fig. 6) bought from the Young Industries, Inc. of Murray, Pennsylvania. The rotary valve is driven by a ½ hp. variable speed motor.

At first, the rotary valve was installed directly under the dust bin and the valve performed well for two experiments. During the third run, the valve stopped feeding even though it was turning. The valve later was dismantled which revealed that the metal to metal seal between the valve rotor and body was scored by abrasives. Since the downstream side of the valve is at a higher pressure than the upstream, air was escaping past the valve and supporting the dust thus preventing the dust from dropping in the rotating valve.

A 2" ball valve and a small cone vessel were connected in series between the dust bin and rotary valve. The 2" ball valve is opened to allow some dust from the dust bin to drop into the cone vessel and is then close to make the cone vessel air tight. The stirring rod is used to push the dust down through the ball valve when required. A ¼" line is connected between the downstream and upstream side of the rotary valve to ensure the pressures are equal. The dust charge in the conical vessel is sufficient to provide approximately 10 to 15 min. of dust recycling time. The dust can either feed to the bed annulus (Fig. 6) or mixed with fresh coal and fed into the bed spout.

3. EXPERIMENTAL GASIFICATION RESULTS

3.1 Analysis of Prior Air-Steam Gasification Results

During the preparation of Reference 4, which was presented at the 32nd Canadian Chemical Engineering Conference in October 1982 a re-examination of some of the previously reported data was made. This was done to examine differences between spouted and fluidized bed operation of the gasifier, and used coals described in Table 1.

For a given coal and mode of operation, two prime variables are used in the gasification experiments - the coal feed rate, and the air rate. The steam rate may also be adjusted, although the lower rank coals require no steam, because of their contained water. If the air rate is set at some multiple of the minimum spouting velocity of the inerts in the bed and the coal feed rate is varied, the gas heating value goes through a broad maximum as shown in Figure 7. An increase in coal feed rate beyond a critical value causes the reactor to cool, and the coal to accummulate. Eventually the reactor will choke. If the air rate is changed, the heating value curve in Figure 7 will shift. This is shown indirectly in Figure 8 where at a fixed coal feed rate, the heating value decreases linearly with increases in the air to coal mass ratio.

The gas yield is also a function of both air and coal rates, however its behaviour is more simply described. Figure 7 shows that at a constant air rate, even where the heating value goes through a maximum, the gas yield decreases in approximately a linear manner with the increase in coal feed rate. Figure 8 shows that over the range of interest a linear increase in gas yield results from increases in the air/coal ratio.

Since the air/coal ratio is the most common variable used in comparing gasification conditions, most results are presented in plots versus this ratio. The coal feed rate has been accounted for in an empirical manner suggested by the plots of Figures 7 and 8.

Examination of the experimental data showed that plots of heating value versus air/coal ratio showed considerable scatter due to differences in coal feed rate, however the heating values - feed rate product yielded a smooth curve when plotted versus air/coal ratio. Figure 9 shows plots of the heating value - feed rate product and the gas yield/feed rate ratio versus air/coal ratio for all runs taken over three years with Forestburg coals of comparable feed size. A straight line has been fitted to this data.

$$G.H.V. = (221.1 - 29.8\ m_A/m_C)/m_C$$

At a fixed air rate, this equation predicts a maximum in the heating value with coal feed rate, and at fixed coal rate, a linear decrease in heating value with increasing air/coal ratio - the heating value characteristics illustrated in Figure 7 and 8. Within the scatter of the heating value data there appears to be no significant difference between fluidized and spouted bed operation, nor among three different shipments of Forestburg coal tested over the three year period. Data points for smaller and larger coal particle sizes, not shown, appeared below and above the curve respectively. The gas yield plot also shows no apparent difference between spouted bed and fluidized bed modes with coal shipment C. It appears that slightly lower yields were achieved with coal shipment B, and excluding this data a least squares fit yielded

$$G.Y. = m_C(0.062\ m_A/m_C - 0.077)$$

Thus the gas yield increases with the air/coal ratio at constant coal feed rate, and at constant air rate decreases as the coal rate increases as shown in Figures 7 and 8. Data for all experimental runs are given in Appendix B.

Some runs were done in FY 82-83 to complete the picture for Roselyn coal. Figure 11 shows similar heating value and gas yield plots for this coal, which is quite similar in analysis to Forestburg. Here a significant improvement in heating value is apparent when the coal is gasified in a fluidized rather than a spouted bed. Gas yields also appear slightly higher for fluidized bed operation. Parameters of the equations corresponding to (1) and (2) fitted to the data are given in Table 2. The reason for the better results under fluidizing conditions is not immediately clear and some possible contributing factors will be discussed below. Heating values are below those from Forestburg coal which is not surprising as Roselyn coal has a slightly lower volatile content and would tend to be less reactive. Gas yields from both coals are virtually identical. This was the only coal that showed an apparent advantage under the fluidized bed mode compared to the spouted bed mode.

The product of the gas heating value and yield divided by the heating value of the coal fed represents the cold gas thermal efficiency

$$\eta_{cold} = G.H.V. * G.Y./C.H.V.$$

With the simple form of equations used to correlate the experimental data:

$$G.H.V.\ m_C = a - b\ (m_A/m_C) \qquad (1)$$

$$G.Y./m_C = e + f\ (m_A/m_C) \qquad (2)$$

it can easily be shown that the maximum cold gas efficiency is given by

$$\eta_{cold,max} = (ae + (af-be)^2\ /4bf)/C.H.V.$$

and occurs at the optimum air/coal ratio

$$(m_A/m_C)_{opt} = (af-be)/2bf$$

From a practical point of view this optimum must be constrained by the requirement that the gas have a heating value sufficiently high to maintain combustion, $G.H.V._*$ Thus

$$G.H.V._{opt} = (a-b(af-be)/2bf)/m_{C\ opt}$$

$$> G.H.V._*$$

The value of $G.H.V._*$ will dictate the coal feed rate to the gasifier. Optimum air/coal ratios calculated from the fitted equations for the various coals are

given in Table 2. The corresponding feed rates were calculated assuming a value of $G.H.V._{*} = 3.73\ MJ/m^3$. The low values of the maximum cold gas efficiency particularly for Sukunka coal, reflect the large carbon losses at the cyclones. It has been shown (3) that recycling this char will improve the efficiency. Increasing the coal particle size was shown previously (2) to improve heating values significantly in the spouted bed, thus the values of Table 2 do not reflect maximum values of efficiency achievable in the gasifier.

3.2 Air-Steam Gasification of Balmer Rejects (Major Task 6)

Balmer rejects contain roughly 54% ash (Table 1). The focus for the attempt to gasify the rejects came from discussions with G. Sirianni, and suggestions from B.C. Coal. The rejects are a waste product currently being stockpiled.

All results for this coal product must be considered preliminary because their high-ash level, meant that ash at about 40 kg/h was entering the gasifier. The unit was not equipped to handle these ash levels, as there was no bed overflow. Thus an accummulation of ash occured, and the bed level rose significantly during the experiments. The temperatures and gas heating values were relatively constant however. Some results are listed in Table 3 and results for both spouted and fluidized beds plotted after the form of equations (1) and (2) in Figure 11. Straight lines on the plots are not observed for this coal.

Gas heating values for this reject feed are quite low, the poorest achieved for any coal yet attempted. This is not surprising as the bituminous nature of the parent coal which suggests a low reactivity is coupled with the very high ash level. Nonetheless, conditions could perhaps be established such as air pre-heat to make-up the thermal penality and ash overflow which would permit the bed to operate steadily where a stronger gas could be produced. There seems to be a small difference between fluidized and spouted modes for this coal, with the benefit of slightly higher gas heating values for the spouted bed. However results are scattered somewhat.

3.3 Air-Steam Gasification of Balmer Coal (Major Task 1)

Balmer Coal is a moderately caking bituminous coal from the Crowsnest coal field (Table 1), supplied by B.C. Coal. It has been gasified in air/steam mixtures (Table 4). As found for other bituminous coals, the heating values are somewhat low - 3.1-3.4 MJ/m^3 suggesting a low reactivity. However the bed temperature were below 890°C. Results appear roughly comparable to Sukunka coal (Fig. 12) which is also bituminous. Extensive tests were run with Balmer coal in oxygen-steam mixtures as is described in section 3.5.

3.4 Effect of Char Recycle (Major Task 2)

A series of runs was made with the char recycle system described in section 2 using Forestburg coal (Table 5). Results for a range of char rates at constant

fresh coal feed rate of 32.5 kg MAF coal/hr are shown in Figure 13. In these tests the air/fresh coal ratio was held constant. Thus as the char recycle rate increased the actual O_2/C ratio in the gasifier decreased. The effect is similar to a net increase in feed rate of a lower reactivity coal, and the behaviour of Fig. 1 may be expected. The general trends of Fig. 13 with increasing char recycle rate are a very slight increase in gas heating value, and a marked gain in efficiency as shown from both the heating factor GHV x GY plot and the carbon conversion plot. Run 72 had an inordinately low bed temperature and thus does not follow the general trends. Run 66, at the highest recycle rate showed a higher bed temperature, and higher CO_2 content in the off-gas than did the others.

Other run-by-run comparisons with and without recycle can be made from data in Tables 4 and 5. For char recycling rate of about 5 kg/h for Balmer coal no real benefit could be found, however temperatures will change upon recycle, if the oxygen/carbon ratio in the gasifier is not increased. Run 84A and 84B (Table 5) illustrate this. When after 2 hours' operation with recycle at 7.6 kg/h the char recycle was stopped, the bed temperature rose by 55°C. The heating value of the gas increased slightly, however the gas yield/kg fresh coal decreased. The heating factor GHV * GY, MJ/kg coal rose by only 1% but the carbon conversion increased markedly.

3.5 Oxygen-Steam Gasification (Major Tasks 4 and 5)

The operation of the spouted bed gasifier on oxygen was the major change from previous operation which was undertaken in Fiscal Year 1982-83. Runs were done primarily with Balmer coal and with char recycle, although some experiments were done with Forestburg coal and some experiments without char recycle (Table 6). In the initial oxygen runs (Runs 76-80), air was still used to convey the coal into the gasifier and the associated nitrogen limited the achievable gas heating values. In later experiments a smaller diameter conveying line was tried using air for conveying, however blockage became a problem. Then the system was re-piped to permit oxygen or oxygen-air mixtures to be used for conveying. In all these tests the steam rates were much higher than in the air tests, and the combination of steam and conveying air was used to govern the temperature at a given oxygen/(coal + char) ratio. Figure 14 shows some results. From these graphs several comparisons can be made:

(i) Comparing 76b and 78 at fixed oxygen/(coal + char) ratio, we see the effect of increased fresh coal feed rate raising the heating value of Run 76b (M_c = 43.1 kg/h) to 7.9 MJ/m^3 versus 6.1 MJ/m^3 for Run 78 (M_c = 37.1 kg/h, M_{ch} = 4.84 kg/h). N_2, T, S constant.

(ii) Comparing Runs 80 & 82, we see that reducing the nitrogen flow from 30 to 10 kg/h increases temperature, and gas heating value from 5.3 to 8.3 MJ/m^3.

(iii) Comparing Runs 83 & 82, no decrease heating value is produced, although the oxygen (coal + char) ratio is slightly increased because the effect is offset by a decrease in average bed temperature.

(iv) Given relatively minor changes in O_2/(coal + char) from say 0.5 to 0.8, the gas heating value pretty well reflects the values of the bed temperature (the latter being inversely dependent on N_2 flow).

Figure 15 shows that as the nitrogen flow is increased the heating value drops in a near-linear manner. The intercept at zero nitrogen input agrees roughly with the heating values calculated from the compositions of Runs 82 and 83 corrected to a nitrogen-free basis.

The H_2/CO molar ratios are in the range 0.8 to 1.27 and vary with temperature and the steam/coal ratio. A comparison with results published on other gasifiers is made in section 6.

3.6 In-Bed Gas Sampling (Major Task 8)

The objects of this task are to obtain radial and axial gas composition and bed temperatures during gasification. Graduate student R. Srinivasan began this work but was asked to withdraw from the university in August 1982. Experiments were tried with a single ½-in. O.D. stainless steel probe using a suction pump to draw the gas sample from the bed. Problems were encountered with air contamination, and blocking of the probe with solids. Purging with nitrogen did not solve the problem. Melting of the probe tip also occurred. After several trials with different tips the probe tip design shown in Fig. 16a was adopted.

Z. Haji-Sulaiman, M.A.Sc. student has carried on this work. The proposed gas sampling line is given in Figure 16.

Samples from the gasifier will first be cooled down by passing through a coil placed in a bucket of ice, and then through a filter to remove the entrained coal fines and some water condensate. The remaining water in the stream will be removed in a knockout drum immediately after the filter, before the sample is finally collected. The volumetric flowrate of gas sample drawn out of the gasifier is indicated by the gas flowrator.

Samples will be taken from 6 different levels and at each level, 3 different radial positions in the annulus and 1 in the spout. All of these locations are within the bed of height of about 0.6 m. In all 24 independent samples will be analysed for each set of operating conditions.

It is also necessary to obtain the temperature readings at these 24 locations. Based on the assumption that the bed is symmetrical about the central axis, thermocouple readings will be taken at locations on the same levels and distances from the gasifier wall but opposite the sampling probes.

To ensure that representative sample at any particular lcoation is obtained, an isokinetic sampling procedure is to be adopted. Longitudinal variation of gas velocity in the annulus will be determined using velocity distribution correlations available. Hence direct measurement is not necessary.

Analysis of samples will be made at the end of each run using the gas chromatograph.

3.7 Sulphur Capture (Major Task 7)

This task was not completed in Fiscal Year 1982-83 due to a number of problems. The aims were to reduce the sulphur content of the off-gas, previously measured at about 230 ppm for Forestburg and Sukunka coals by injection of limestone or dolomite into the bed. The gas chromatograph used for routine analysis will not detect sulphur compounds. A chromatograph equipped with a flame photometric detector used previously for sulphur compounds could not be kept operating in a reliable manner, and so it was decided to use wet chemical methods to analyze for sulphur compounds. A gas sampling train (Fig. 17) was set up and used for Forestburg, Run 77.

The sampling system consisted of a gas filter, three glass impingers, air pump, rotameter, and wet-gas-meter. Flue gas coming from the reactor was delivered by the air pump through the gas filter to three glass impingers in series. The flow rate was simply regulated by the rotameter at 1.5 litres per minute. Additionally, the wet-gas-meter was connected to the rotameter to monitor the total amount of flue gas passing through the three glass-impingers over a period of time. 0.1 M NaOH solution with 3% H_2O_2 solution was used for SO_2 absorption, and 0.1 M lead acetate solution was employed for trapping H_2S.

SO_2 test

The sample solution obtained from the glass impingers was adjusted to pH between 2.5 to 4.0 with perchloric acid, 2 drops of 2% thorin indicator was added and titrated with barium perchlorate solution, taking the change from yellow-orange to pink as the end point. No obvious colour change was found from the titrations, therefore no SO_2 was absorbed by the reagent solution during experimentation.

H_2S test

The dark-brown suspension was collected and washed with dilute nitric acid to remove any basic lead acetate, which might be present in the solution owing to high concentration of CO_2 in the flue gas stream. The dark brown precipitate was then collected over micropore filtering device ($\sim$ 0.45 μm). After washing with distilled water a few times, the dark brown precipitate was put in the oven overnight at 110°C to dry any remaining moisture. The nominal weight of H_2S was found to be 1.5 mg in a total collected flue gas volume of 71 litres, this is equivalent to 1 ppm of H_2S by volume.

The reliability of this single low value for H_2S is unknown. An analysis of the ash portion of the cyclone char showed CaO = 6.98 wt%, MgO = 1.75 wt%, while the char itself analyzed 0.49% S. There is sufficient CaO and MgO to bind all the sulphur as CaS and MgS. However calculations indicate that roughly 18% of the inlet sulphur reports with the char which is about the same value found previously (2). Some sulphur may well react with the bed material, but the majority is thought to escape with the gas.

The sulphur analysis of the gas by wet methods is frought with potential errors. A new Gas Chromatograph with a suitable detector has just been installed in the department for another project and some further chromatographic measurements will be made. At such time Major Task 7 can be completed.

4. MODELLING OF THE GASIFIER (Major Task 9)

In last year's report (3) results from a kinetic-free model were shown. In the current year some progress has been made towards a kinetic model based on the approach of Amundsen and coworkers (5). This work was interrupted as described in section 6, by the resignation of Dr. C.J. Lim from paid participation in the project, and has been recently picked up by graduate student Z. Haji-Sulaiman and the graduate students in a Coal Utilization course given by the principal investigator. The work to date has been on a back-mixed reactor kinetic model for char gasification. The approach of the model is to consider a very shallow combustion zone at the bottom of the reactor, with a gasification zone above (Figure 18). All oxygen in the blast is consumed in the combustion zone, and the only reactions in the gasification zone are:

$$H_2O + C \rightarrow CO + H_2 \qquad R - 1$$

$$H_2O + CO \rightarrow CO_2 + H_2 \qquad R - 2$$

Thus CH_4 formation by the C - H_2 reaction is ignored. The kinetics of Eucker are assumed for reaction R - 1

$$g_1 (T,P_i) = k_{01} (P_{H_2O} - P_{CO}P_{H_2}/K_1)$$

$$= k_{01} P_T(x_1 - x_4x_5P/K_1)$$

Eucker reports

$$k_{01} = 625.83 \exp(-21,200 / T)$$

$$K_1 = 10^{(7.49 - 7070/T)}$$

The water-gas reaction R - 2 links R - 1 to the CO_2 - C reaction:

$$K_{wg} = 10^{1.6495 - 1855.6/T} = x_5x_3/x_1x_4$$

The carbon-balance gives the mean carbon-conversion

$$\bar{X} = 1 - \frac{1}{1 + g_1\tau_s}\left[1 - y_2^o/\alpha\right]$$

where τ_s = residence time $\qquad$ y_2^o = mole fraction oxygen in blast

and α = moles C fed/mole gas fed

Elemental balances are set up for H, C, O and N.

The approach used is as follows. For a given temperature, residence time and input blast conditions,guess a value for g_1, calculate $\bar{X}$, calculate the mole fractions of each gaseous species. Then re-calculate g_1 and compare with the assumed value. Iteration is continued until the difference between successive g_1 values is small. A typical result is shown as Fig. 19, using Eucker's kinetics.

The model is now being applied to the gasifier data. The mean residence time, temperature and blast composition are first determined, and the reaction rate is an unknown. This model applies only to the char/gas reaction, and the volatiles will be handled as described in a previous report (3). It is intended to continue working on a two-zone model which reflects the fluid mechanics of spouting, to compare with the backmixed model described above.

5. COMPARISON OF SPOUTED BED AND COMMERCIAL MEDIUM CALORIFIC VALUE GASIFIERS

A comparison of some performance data for the spouted bed gasifier with fully developed commercial and semi-commercial gasifiers is shown in Table 6. For the spouted bed the gas compositions have been corrected for the nitrogen content of the transport air. Thus about 10% N_2 for Balmer and 16% N_2 for Forestburg runs have been subtracted from the off-gases. Several points are of note:

(i) Gas compositions are similar to the Lurgi dry ash moving bed in terms of H_2/CO ratios, and CH_4 contents are between those of a fluidized and a moving bed gasifier as might be expected.

(ii) CO_2 levels are higher than commercial gasifiers. No heat recovery or pre-heating is done for the spouted bed.

(iii) Gross calorific values of gases are about 20% lower for the spouted bed pilot plant than for commercial units. This is also reflected in the CO_2 levels and is no doubt related to poorer thermal efficiencies of pilot scale units.

It would appear that if the spouted bed could be run with optimal particle size coal, and with better heat recovery and lower heat losses, the performance would at least match those of commercial units.

6. PERSONNEL

At the beginning of the year 1982-83, the project was manned by the Principal Investigator and

Dr. C. B. Prakash	Research Engineer in charge
Mr. G. Cheng	Junior Engineer
Dr. C.J. Lim	Research Associate - part time (24 days)
R. Srinivasan	Graduate Student

Dr. Prakash joined CANMET in June, and was replaced in mid-July by Mr. G. Cheng. S. Low was hired into the Junior Engineer's position in August. Dr. C.J. Lim joined the UBC faculty in July and has not been able to carry out the modeling work, and has not been replaced. R. Srinivasan was asked to leave the university and was replaced in September by Z. Haji-Sulaiman. These changes have made it more difficult to complete the Major Tasks in the allotted time.

7. PAPERS AND CONFERENCES

A.P. Watkinson presented a paper (4) at the 32nd Canadian Chemical Engineering Conference in Vancouver, October 1982. This paper has been accepted for publication

in the Canadian Journal of Chemical Engineering.

Mr. Cheng and Mr. Low attended the Coal Sessions of the above-mentioned conference.

8. BUDGET

Table 7 summarizes the money committed under the project up to its extended date of completion April 30, 1983. The extension of one-month was requested because of delays in construction and time lost due to changes in personnel (Section 6). Generally the project was kept under budget, with larger equipment and supplies expenditures than budgeted, and smaller total salary, computer and travel expenditures than budgeted.

9. CONSULTATION WITH S.N.C. (Objective 4)

During the year some discussions were held with personnel from S.N.C. Inc. concerning the scale-up and design of a pilot unit. Comments and suggestions were made concerning their models for scale-up, and an assessment of the final report from this study was sent to CANMET.

10. CONCLUSIONS AND FUTURE WORK

1. A logical method was developed to analyze differences among coals and reactor operating conditions, based on observed variations in heating values and gas yields with independent variables of air/coal ratio and coal feed rate.

2. The modified char recycle apparatus permitted trouble-free recycle of cyclone char either to the fresh coal feed line or to the annulus of the spouted bed. The benefits of operating with recycle were demonstrated on Forestburg coal.

3. The spouted bed was adapted to oxygen/steam gasification and achieved results nearly comparable to commercialized units.

4. A new bituminous coal, Balmer, was tested. Trouble-free operation was experienced both in air/steam or oxygen/steam operation.

5. A washery reject stream Balmer Rejects of about 55% ash was gasified in some preliminary tests. A bed overflow system must be added to draw off accumulating ash. Gas heating values were found to be low.

6. The in-bed gas sampling and gasifier mathematical modelling progressed marginally during 1982-83. More work is required to close the sulphur balances and assess the feasibility of limestone or dolomite injections.

7. A proposed work statement for 1983-84 is given in Appendix A.

REFERENCES

1. A.P. Watkinson, S.K. Foong and G. Cheng "Spouted Bed Gasification of Western Canadian Coals - Final Report 1979-1980" DSS Contract 08SU-23440-8-011, May 1980.

2. A.P. Watkinson, S.K. Foong and G. Cheng "Spouted Bed Gasification of Western Canadian Coals - Final Report 1980-81. DSS Contract 08SU-23440-8-011, April 1981.

3. A.P. Watkinson, C.B. Prakash, G. Cheng and C.J. Lim "Spouted Bed Gasification of Western Canadian Coals - Final Report 1981-82 Contract No. OSU81-00074, Energy Research Labs CANMET, May 1982.

4. A.P. Watkinson, G. Cheng and C.B. Prakash "Comparison of Coal Gasification in Fluidized and Spouted Beds" Preprint Proceedings Vol. 1, 32nd Can. Chem. Engn. Conference, Vancouver, B.C. Oct. 3-6 1982, p. 433-442.

5. S. Sundaresan and N.R. Amundson "Studies in Char Gasification - I A Lumped Model" Chem. Eng. Sci. 34, 345 (1979).

NOMENCLATURE

a,b,e,f	Constants in equations (1) and (2)
C.H.V.	Gross heating value of coal MJ/kg MAF coal
G.H.V.	Gross heating value of gas MJ/m^3
G.Y.	Gas yield M^3 (STP)/kg MAF coal
g_1	Reaction rate function
k	Rate constant
K_1, K_{WG}	Equilibrium constants for reaction R-1 and R-2 respectively
M_A	Air mass feed rate kg/hour
M_C	Coal feed rate kg MAF coal/hour
M_{Ch}	Char feed rate kg MAF char/hour
M_S	Total steam rate including H_2O in coal kg/hour
η	Thermal efficiency

ACKNOWLEDGEMENTS

Coals were kindly supplied by B. C. Coal, Luscar Ltd., and Manalta Coal Ltd. Facilities for crushing and screening were provided by B. C. Research.

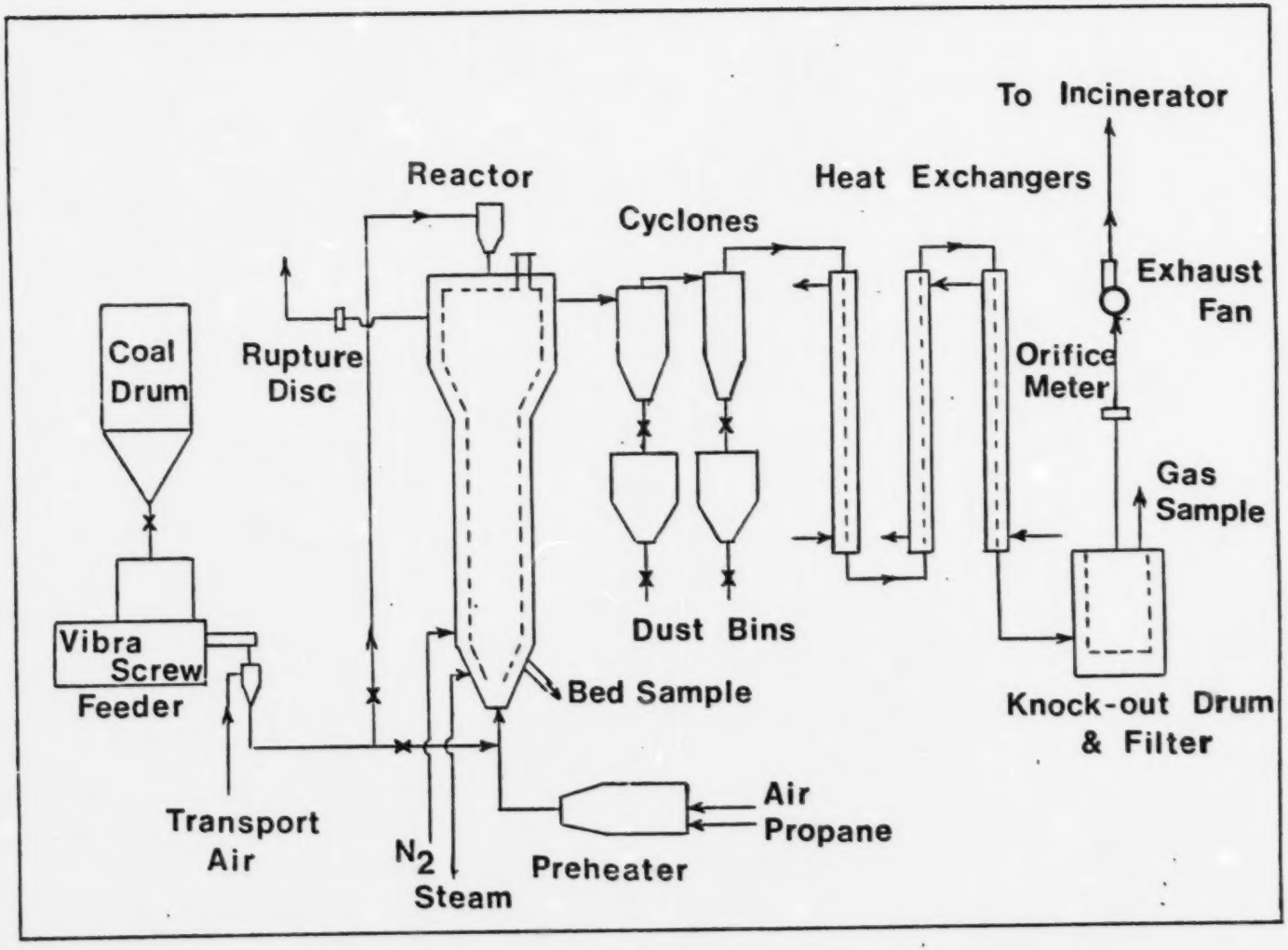

FIGURE 1

Flow Diagram of the Process Development Unit

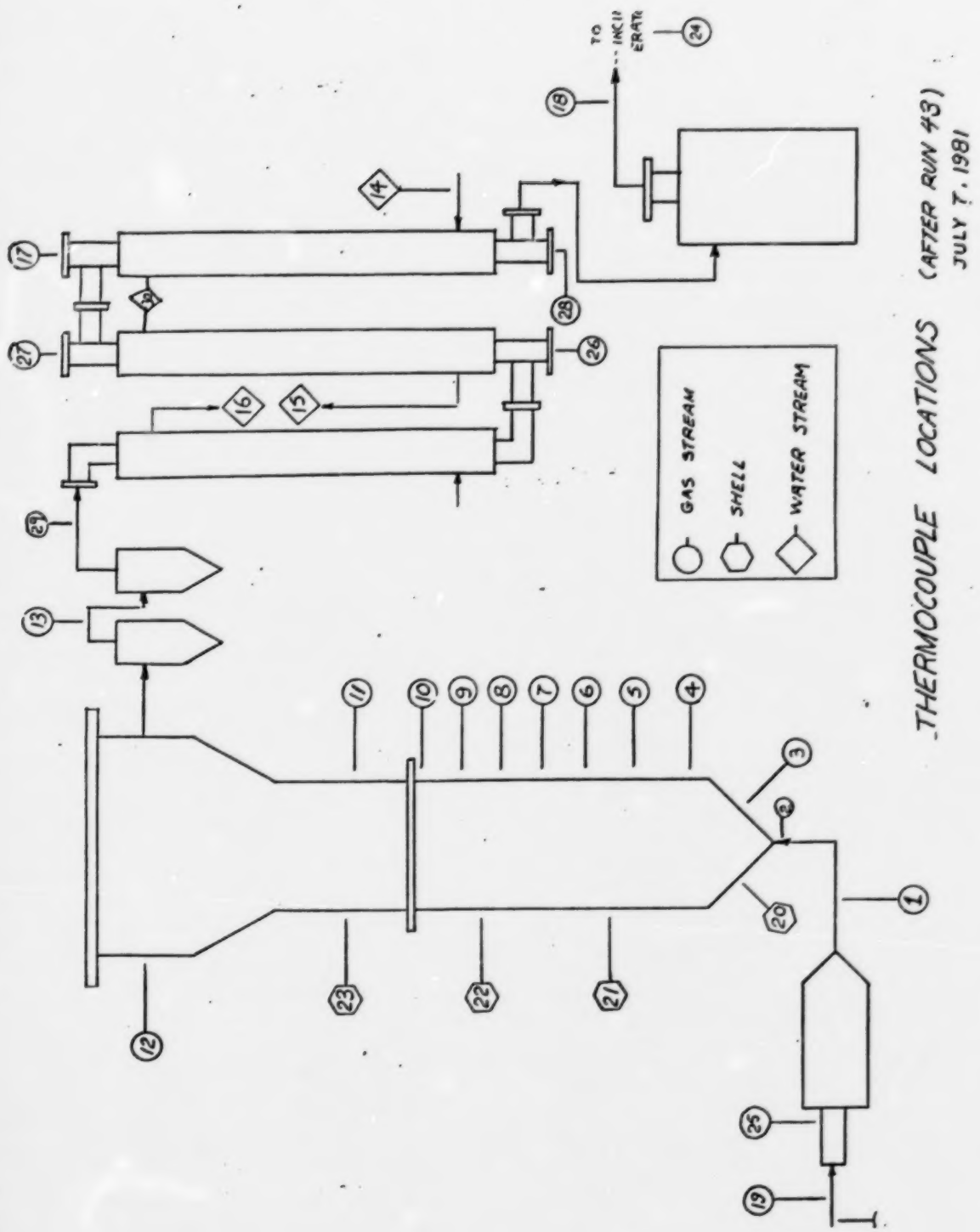

Figure 2. Thermocouple Location Diagram

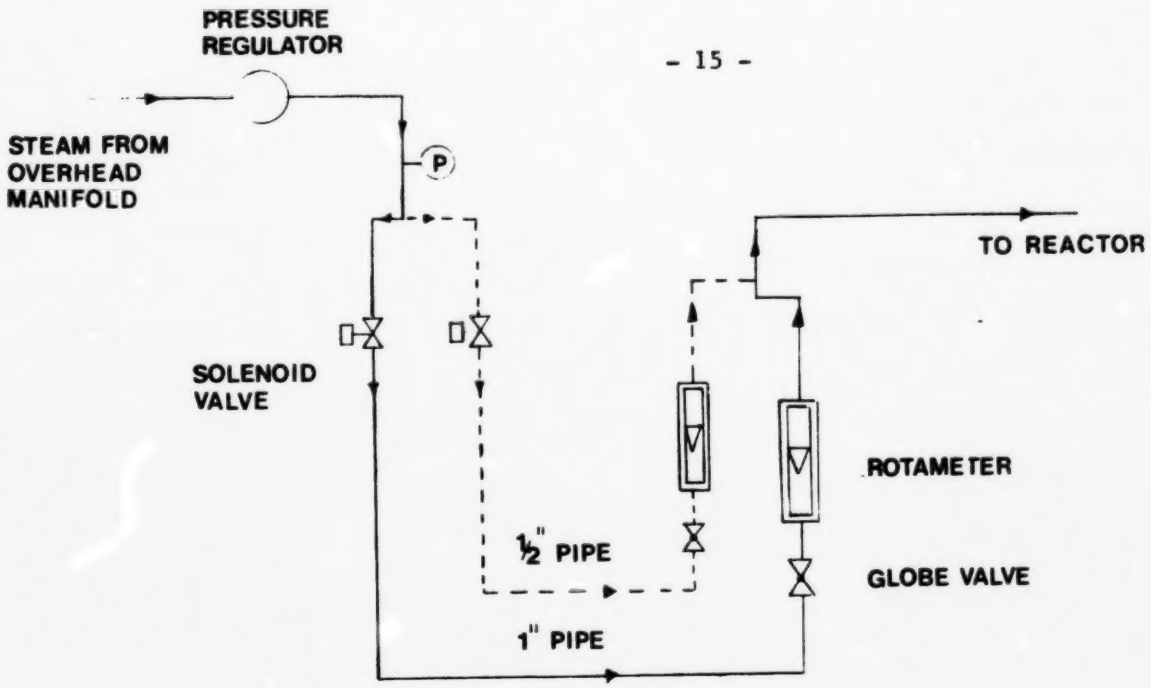

Fig. 3 - Steam Line Modifications

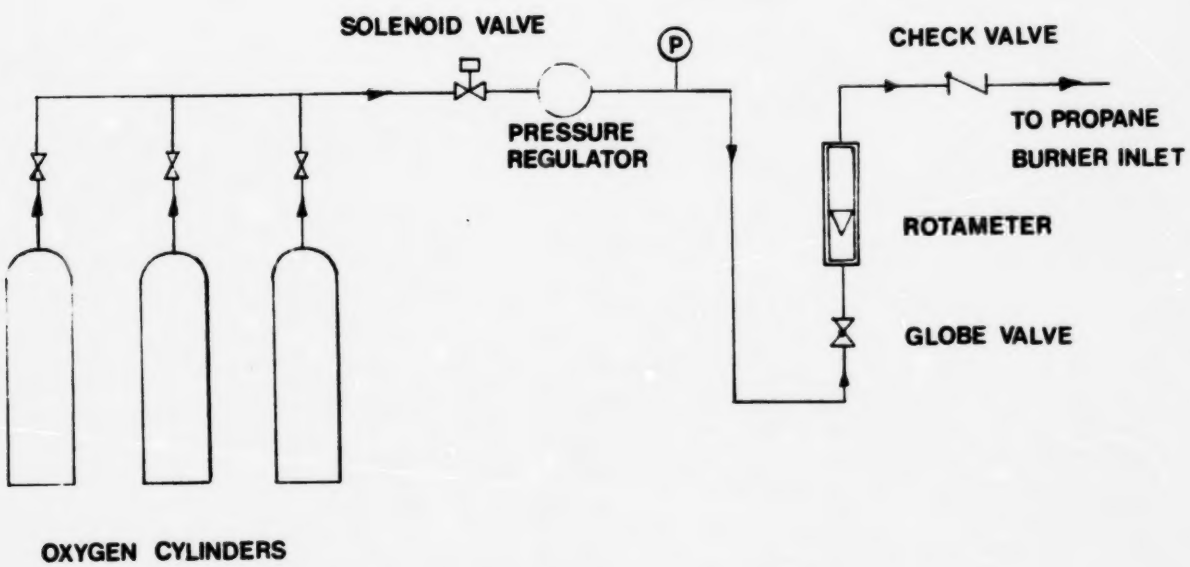

Fig. 4 - Oxygen Installation

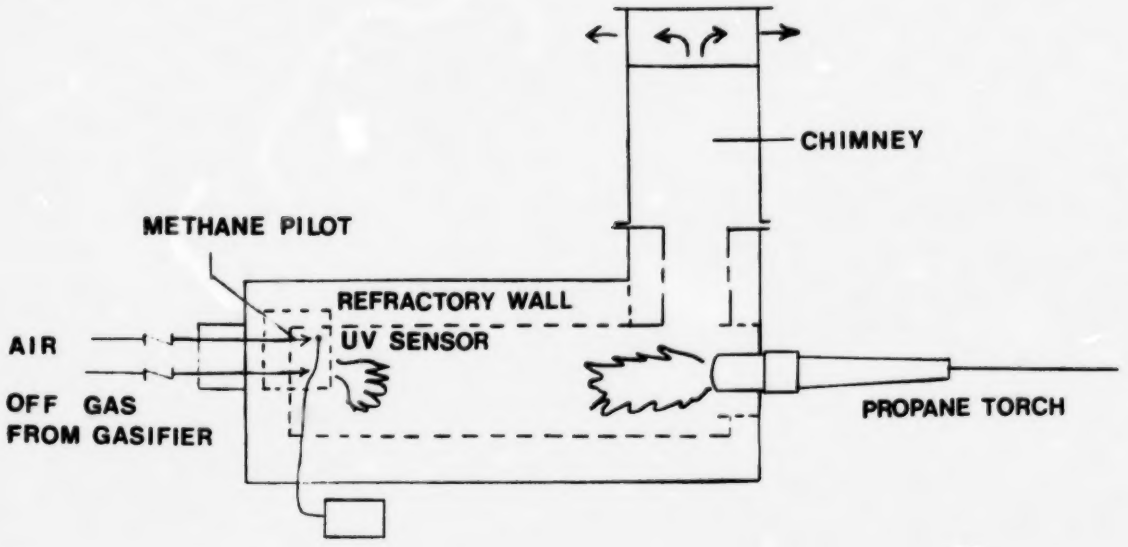

Figure 5 Off-gas Incinerator Schematic

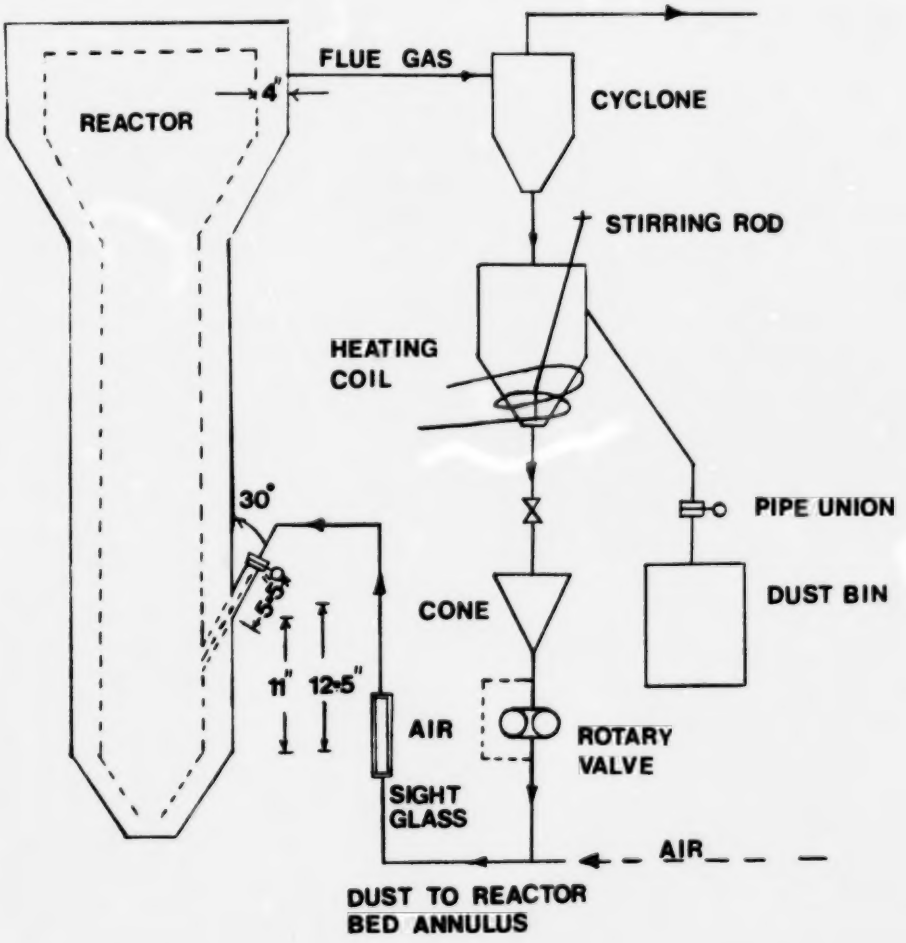

Figure 6 Cyclone Char Recycling System

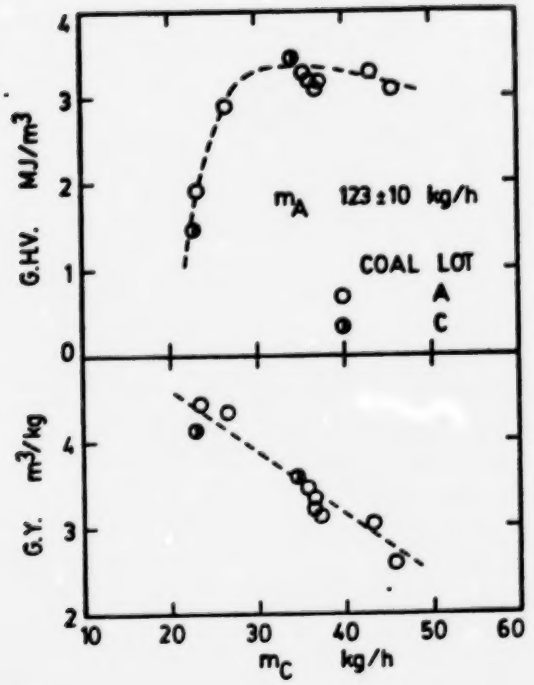

Fig. 7 Effect of coal feed rate (M.A.F. basis) at a constant air rate on gas heating value and gas yield for Sukunka coal in a spouted bed. (Coal shipment A = 1979-80, C = 1981-82)

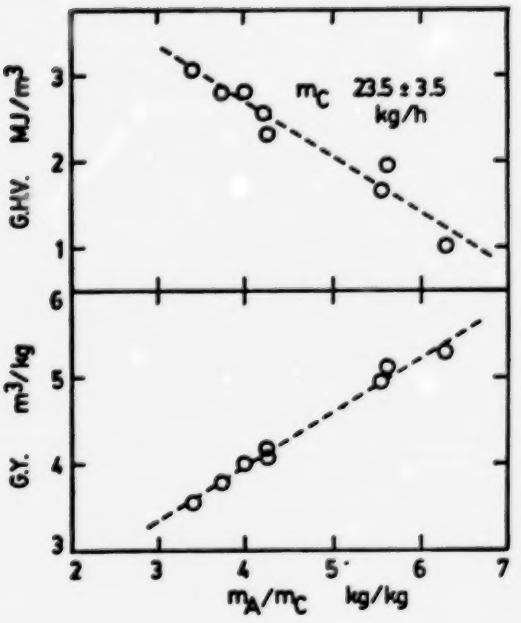

Fig. 8 Effect of air/coal mass ratio (M.A.F. basis) at a constant coal rate on gas heating value and gas yield for Roselyn coal in a spouted bed.

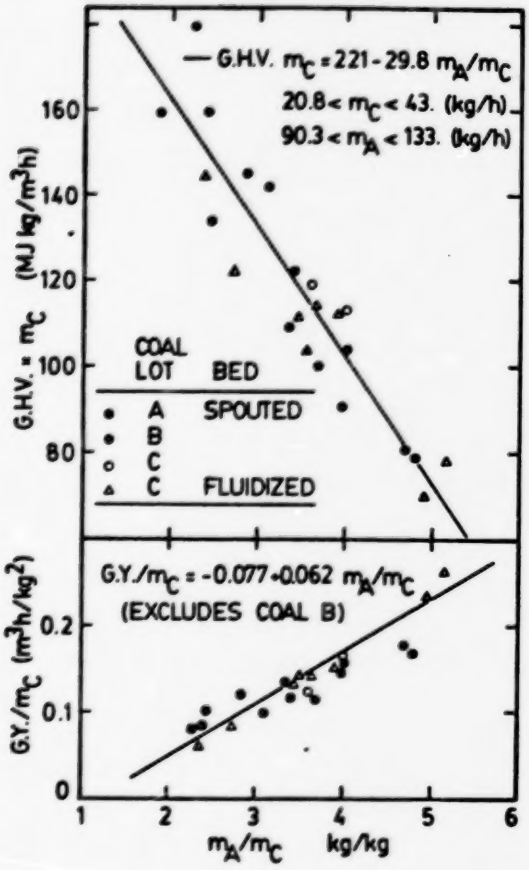

Fig. 9 Gas heating value and yield functions versus air/coal ratio for Forestburg coal in spouted and fludized beds (A = 1979-80, B = 1980-81, C = 1981-82 Coal Shipments)

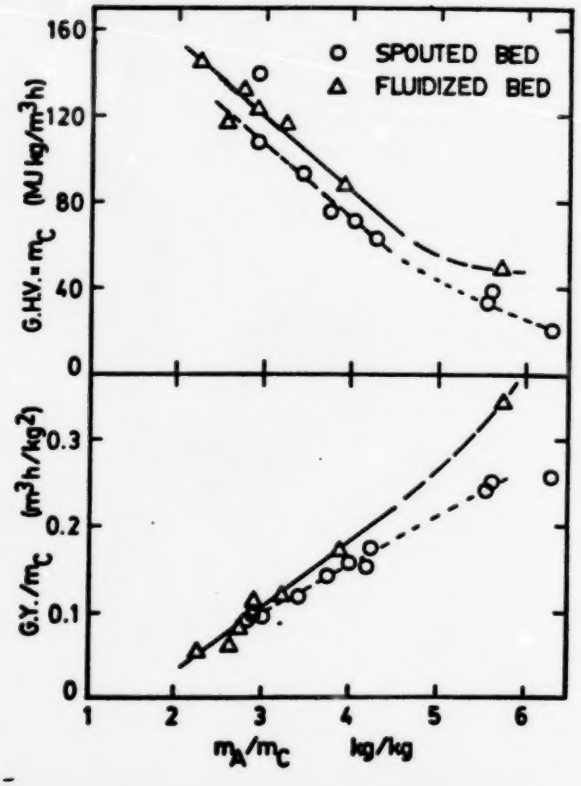

Fig. 10 Gas heating value and yield functions versus air/coal ratio for Roselyn coal in spouted and fluidized beds

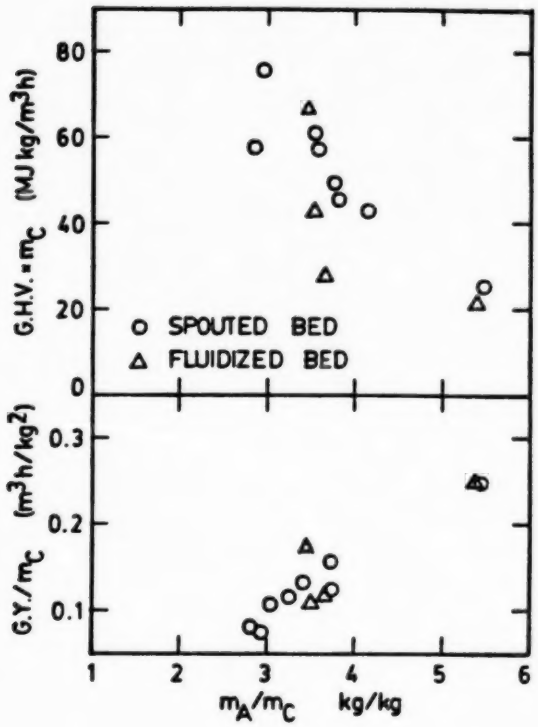

Fig. 11 Gas heating value and yield functions versus air/coal ratio for Balmer Rejects in spouted and fluidized beds

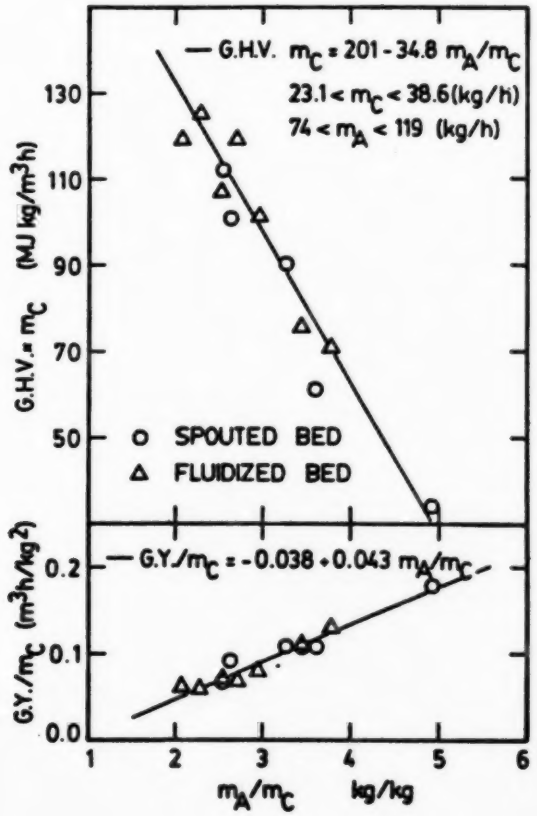

Fig. 12a) Gas heating value and yield functions versus air/coal ratio for Sukunka coal in spouted and fluidized beds

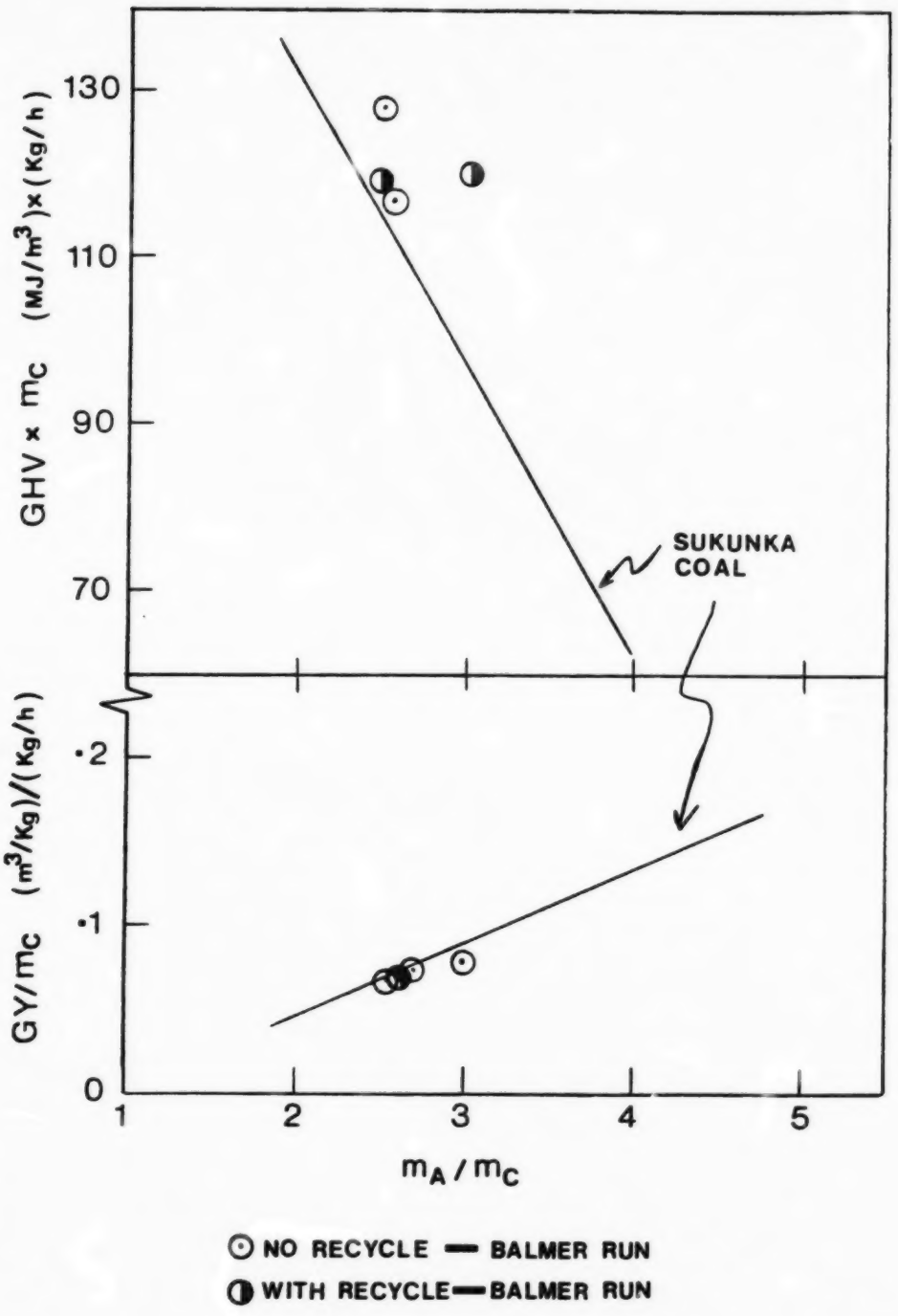

Fig. 12b) Gas Heating Value and Yield Functions versus Air/Coal Ratio for Sukunka and Balmer Coals

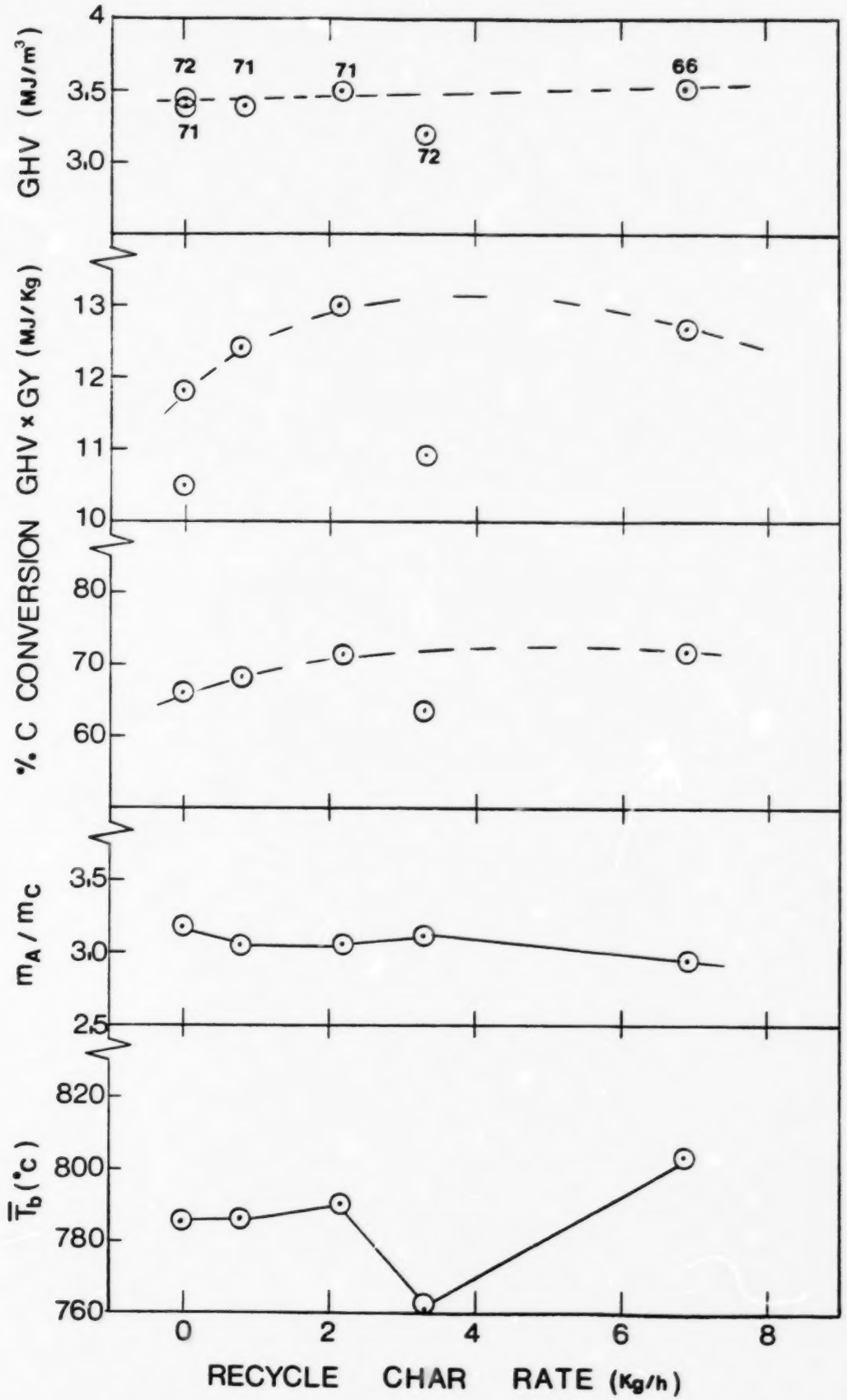

Fig. 13 Effect of Char Recycle Rate - Forestburg Coal
(Numbers refer to Experimental Runs)

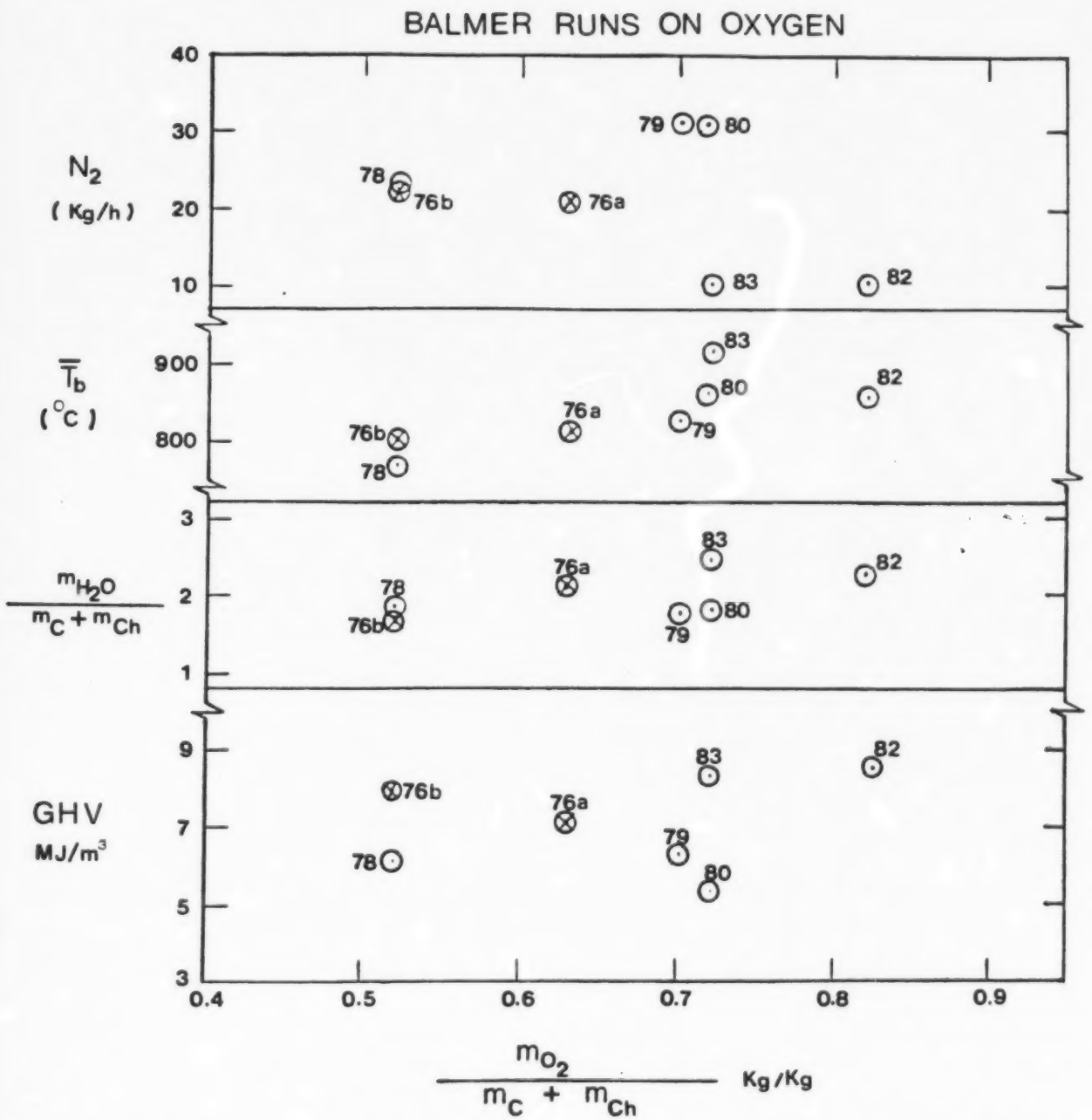

Fig. 14 - Oxygen - steam Gasification - Effect of O_2/Coal Ratio
(Numbers refer to Experimental Runs)

⊙ No Recycle ⊗ Char Recycle

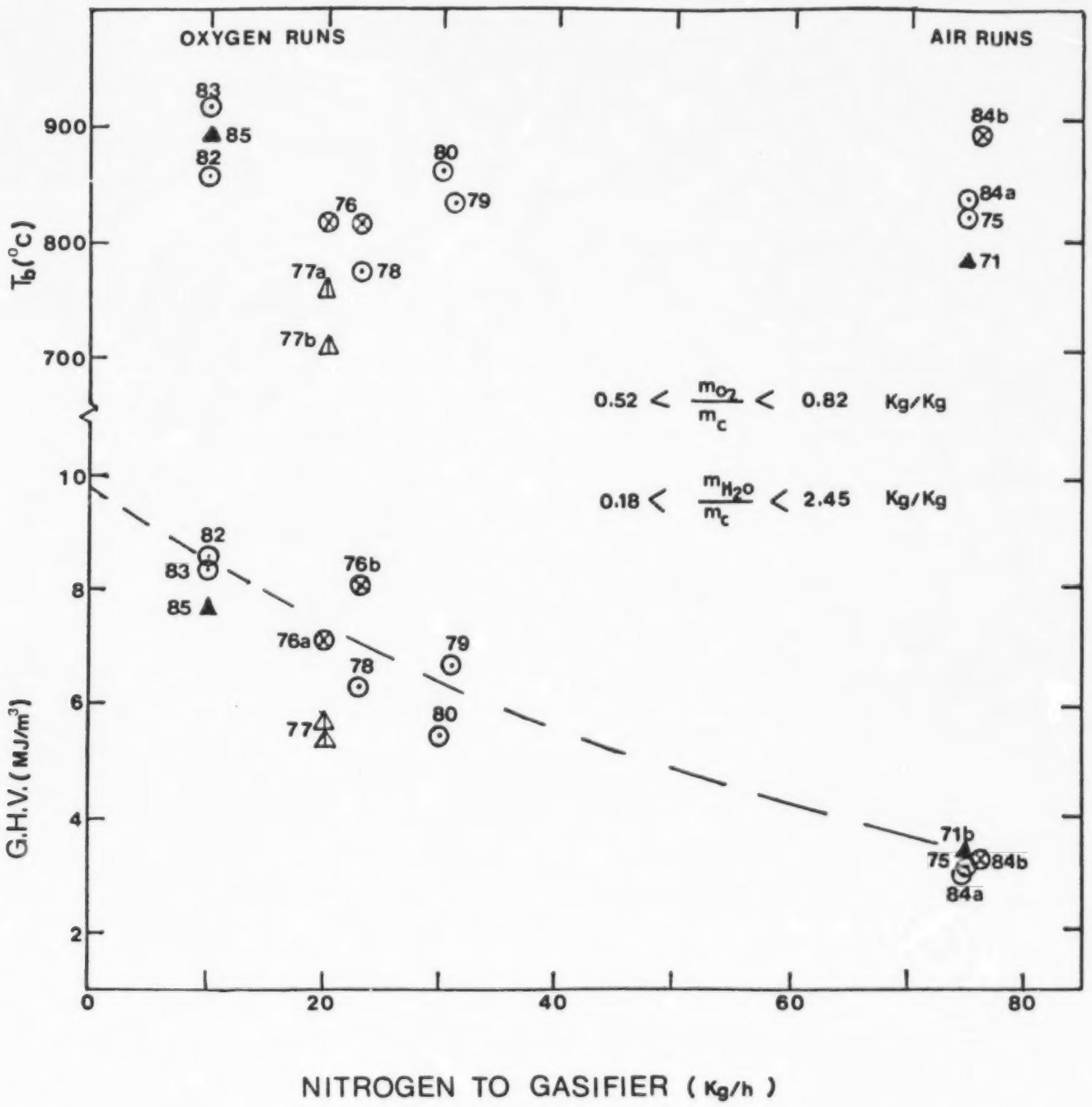

Fig. 15 - Effect of Nitrogen Flowrate on Gas Heating Value and Bed Temperature

⊙ Balmer - No Recycle
⊗ Balmer - Char Recycle
▲ Forestburg - No Recycle
▲ Forestburg - Char Recycle

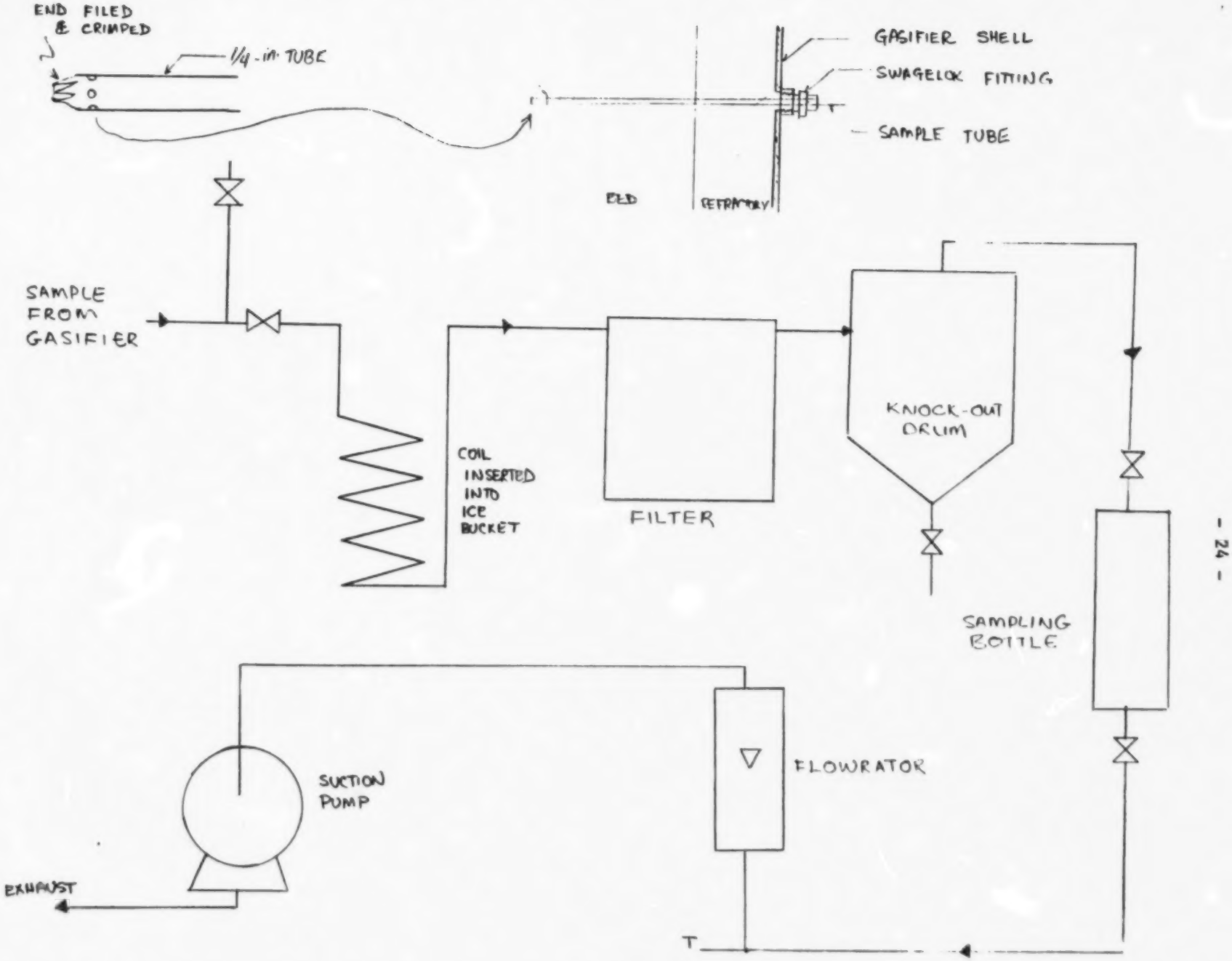

Fig. 16 a) SAMPLE TUBE & MOUNTING b) GAS SAMPLING TRAIN

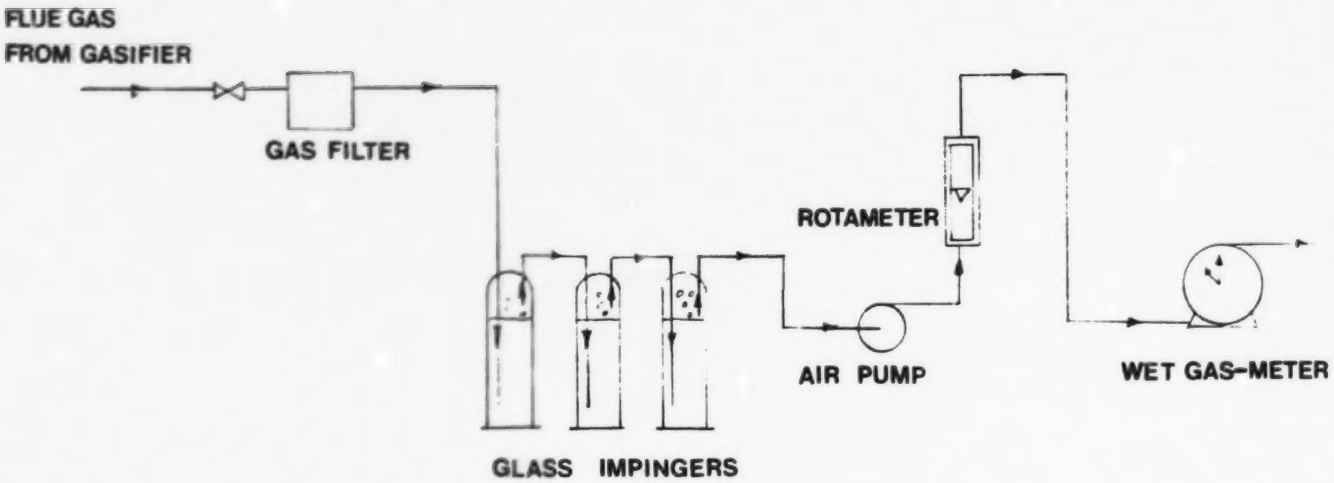

Figure 17 Gas Sampling Train for Sulphur Analysis

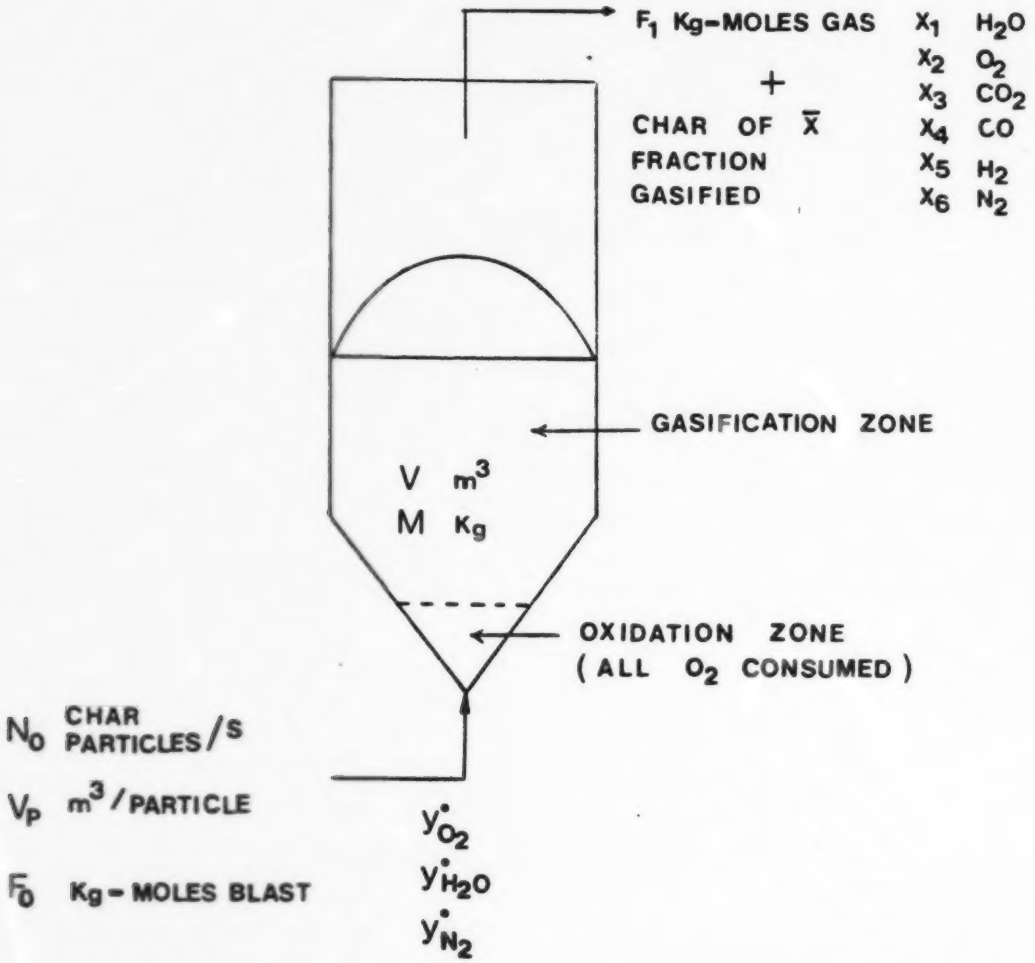

Fig. 18 Schematic Kinetic Model of Backmixed Gasifier

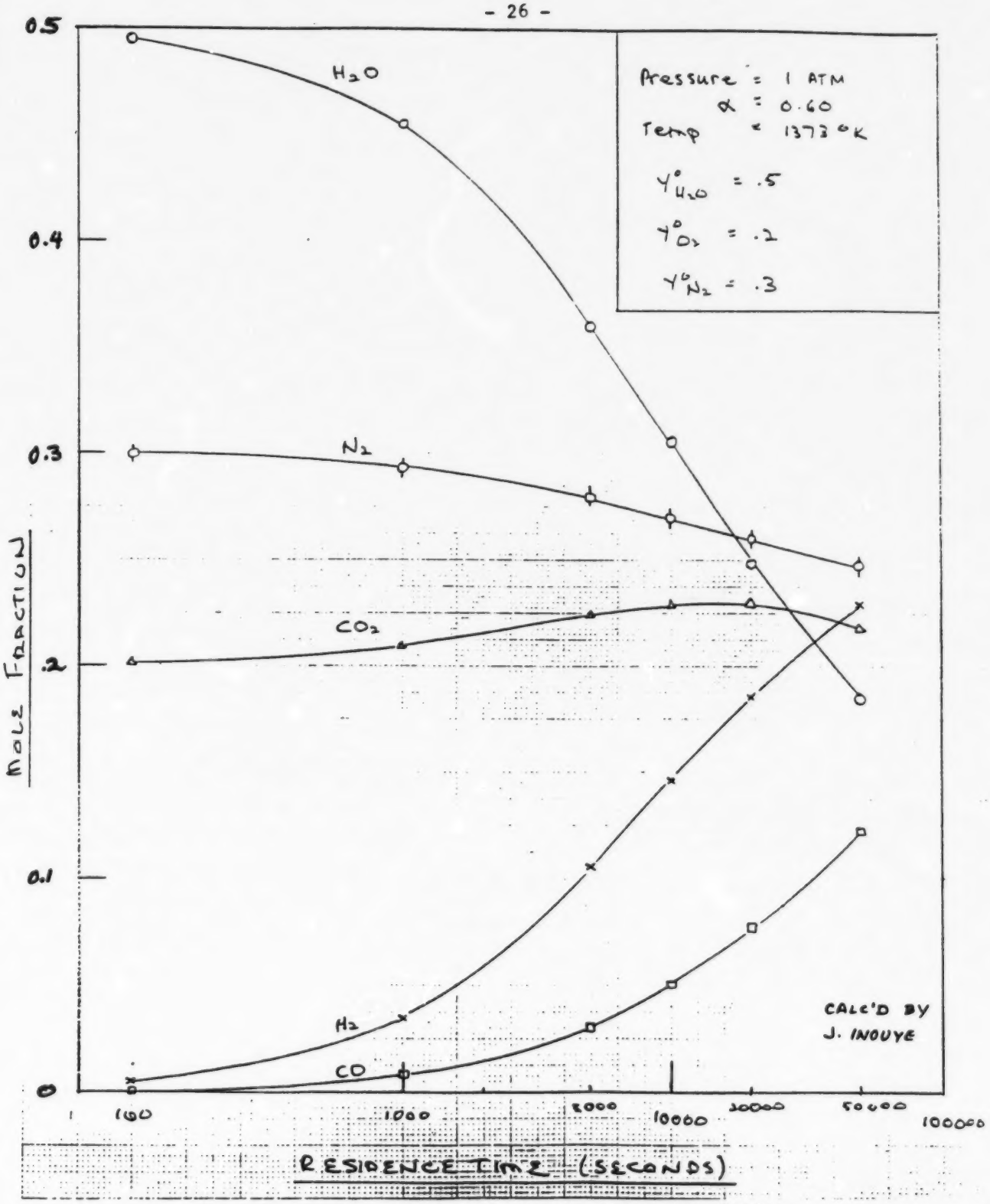

Fig. 19 - Typical Backmixed Kinetic Model Calculations using Amundson Approach and Eucker Kinetics

Table 1 Coal Analysis

	Forestburg	Roselyn	Balmer	Balmer Rejects
1. Ultimate (dry basis)				
% C	61.53	71.34	78.94	38.50
H	3.65	4.44	4.09	2.07
N	1.31	1.43	1.00	0.57
O	15.16	16.65	4.07	5.20
S	0.53	0.69	0.22	0.13
Ash	17.82	5.45	11.68	53.53
2. Proximate (wet basis)				
% Moisture	23.71	27.01	1.47	1.17
Volatile	29.44	29.11	20.07	16.47
Fixed carbon	33.25	39.90	66.95	29.46
Ash	13.60	3.98	11.51	52.90
Calorific value MJ/kg	18.15	20.27	31.24	14.14

Table 2 Fitted Curves and Estimated Optimal Conditions for Various Coals

Coal	Mode of Operation	Value of Constants*				$\eta_{cold,max}$	$(m_A/m_C)_{opt}$	$m_{C_{max}}$ **
		a	b	e	f			(kg/h)
Forestburg	Sp & Fl	221.1	29.8	-0.077	0.062	0.60	4.33	24.7
Roselyn	Sp	210.4	34.9	-0.073	0.057	0.38	3.65	22.3
	Fl	222.0	34.5	-0.122	0.075	0.51	4.03	22.2
Sukunka	Sp & Fl	200.8	34.8	-0.038	0.043	0.23	3.35	22.8

* G.H.V. $= (a - b(m_A/m_C))/m_C$ MJ/m^3

G.Y. $= (e + f(m_A/m_C))m_C$ m^3/kg

** For G.H.V. > 3.73 MJ/m^3

Table 3 Gasification of Balmer Rejects

RUN No	FEEDRATE (kg/h) m_c	m_A	m_s	m_{ch}	m_A/m_c	m_s/m_c	TEMP. °C AVG.	MAX.	GAS. HEAT. VAL. MJ/m³	GAS YIELD m³/kg	GAS ANALYSIS (%MOL) H_2	CO	CO_2	CH_4	N_2	% C CONVERSION
62a	18.5	99.45	6.73		5.38	0.36	904	946	1.38	4.45	4.5	4.6	13.7	0.71	76.5	51.9
62b	29.4	102.4	10.8		3.48	0.36	863	874	2.22	3.03	7.0	6.77	12.42	1.46	72.3	38.5
62c	36.0	104.5	10.5		2.90	0.29	868	880	2.13	2.49	7.5	6.08	11.55	1.30	73.6	29.0
63a	27.5	101.6	10.1		3.69	0.37	907	936	1.82	3.23	6.0	5.98	14.68	0.98	72.4	43.1
63b	29.4	103.9	10.6		3.53	0.36	879	908	1.99	3.20	7.34	5.85	15.99	1.04	69.8	44.9
64a	32.4	90.4	4.79		2.79	0.15	820	872	1.79	2.48	6.98	5.83	15.45	0.63	71.1	33.4
64b	26.4	98.5	9.17		3.73	0.34	899	997	1.75	3.34	6.26	6.84	15.86	0.45	70.6	47.6
64c	24.7	100.5	9.91		4.07	0.40	674	910	1.76	3.68	7.26	7.10	15.73	0.15	69.8	51.9
65a	18.0	95.1	3.66		5.28	0.20	756	807	1.20	4.30	2.62	4.67	13.28	0.84	78.6	49.6
65b	26.6	96.2	5.47		3.62	0.20	709	764	1.06	2.93	2.65	3.49	14.59	0.84	78.5	32.7
65c	28.3	97.5	5.54		3.45	0.20	758	870	1.55	2.90	4.19	4.77	13.99	1.10	75.6	36.4
69*	27.0	99.8	7.28		3.70	0.27	700	706	1.86	3.32	6.53	5.31	17.15	1.12	69.7	48.3
70*	25.0	89.0	1.00		3.56	0.04	663	930	2.72	3.56	7.60	9.64	18.03	1.70	63.0	64.3

* FLUID BED RUNS

m_c = COAL ON M.A.F. BASIS

TABLE 4 Gasification of Forestburg Coal in Oxygen and Air-Steam Mixtures

RUN NO.	FEEDRATE* (kg/h) m_{O_2}	m_C	m_A	m_S	m_{Ch}	m_{ox}/m_C	m_S/m_C	TEMP. °C AVG.	MAX.	GAS. HEAT. VAL. MJ/m³	GAS YIELD m³/kg	GAS ANALYSIS (%MOL) H_2	CO	CO_2	CH_4	N_2	% C CONVERSION
77a	16.70	36.44	27.00	91.8	—	0.63	2.52	761	787	5.50	1.72	30.64	9.17	31.32	1.67	27.20	48.32
77b	17.60	42.45	27.30	88.07	—	0.56	2.07	710	716	5.40	1.62	29.77	8.41	33.54	2.03	26.25	47.89
85	26.80	31.93	12.90	96.68	1.42	0.93	3.03	881	903	7.50	2.35	30.98	22.65	29.71	2.68	13.98	86.89
66		34.08	100.00	18.87	6.87	0.68	0.55	804	822	3.51	3.60	13.44	8.11	25.76	1.13	51.55	71.72
71a		32.23	97.66	10.96	0.81	0.71	0.34	782	786	3.38	3.69	13.43	12.33	15.86	1.53	56.85	68.04
71b		32.23	98.17	10.96	2.08	0.71	0.34	789	792	3.48	4.00	14.42	14.46	14.49	1.46	55.18	71.65
72		31.07	99.99	11.76	3.33	0.75	0.38	759	777	3.20	3.40	14.02	11.83	15.25	1.47	57.42	63.24
73		27.05	96.84	12.71	—	0.83	0.47	792	827	3.87	3.09	13.89	10.20	17.14	0.47	58.30	72.59

* m_C = COAL ON M.A.F BASIS $\quad m_{ox} = m_{O_2} + 0.23\, m_A$ $\quad m_A$ = AIR $\quad m_{O_2}$ = PURE OXYGEN

TABLE 5 Gasification of Balmer Coal

RUN NO.	FEEDRATE* (kg/h) m_{O_2}	m_C	m_A	m_S	m_{Ch}	m_{ox}/m_C	m_S/m_C	TEMP. °C AVG.	MAX.	GAS. HEAT. VAL. MJ/m³	GAS YIELD m³/kg	GAS ANALYSIS (%MOL) H_2	CO	CO_2	CH_4	N_2	% C CONVERSION
76a	15.50	34.97	28.20	75.26	—	0.63	2.15	807	816	7.13	1.47	19.27	24.32	17.52	6.44	32.45	43.00
76b	15.90	43.06	27.00	72.73	—	0.52	1.69	801	809	7.86	1.31	21.47	22.76	19.36	6.69	29.72	32.66
78	15.70	37.06	28.80	77.33	4.84	0.60	2.09	780	829	6.12	1.33	15.72	23.17	20.08	4.46	36.57	35.26
79	19.30	35.23	40.40	72.62	5.58	082	2.06	824	856	6.27	1.95	18.46	19.45	20.45	3.87	37.77	51.00
80	19.10	34.97	39.90	72.57	5.58	0.83	2.08	863	868	5.36	1.88	21.72	17.08	24.41	2.21	34.58	56.98
82	31.80	37.53	13.00	96.45	5.47	0.94	2.57	853	875	8.50	1.72	29.89	28.32	31.16	3.56	7.07	61.93
83	28.70	38.89	13.30	108.50	5.32	0.82	2.79	917	932	8.27	1.68	32.88	26.90	27.80	2.75	9.10	63.09
75		37.58	95.33	6.64	—	0.59	0.18	821	841	3.00	2.43	9.89	10.40	12.76	1.45	65.50	34.14
84b		37.32	99.64	7.46	—	0.62	0.20	890	895	3.22	2.37	12.44	10.34	13.55	1.71	61.95	38.70
81		37.32	111.96	7.84	4.09	0.70	0.21	881	912	3.08	2.82	9.69	10.67	10.78	1.63	67.23	37.02
84a		37.32	97.78	7.46	5.18	0.61	0.20	836	845	3.01	2.50	9.96	9.05	12.87	1.87	66.25	33.63

* m_C = COAL ON M.A.F. BASIS $m_{ox} = m_{O_2} + 0.23\, m_A$ m_A = AIR m_{O_2} = PURE OXYGEN

TABLE 6 Comparison of Oxygen Fired Gasifiers

Gasifier	Winkler	KT	Texaco	Lurgi	Westinghouse PDU	Spouted Bed Pilot	
Gasifier Type	Fluidized Dry Ash	Entrained Slagging	Entrained Slagging	Moving Dry Ash	Fluidized Agglo-Ash	Dry Ash	
Coal	Lignite	Bit.	Bit.	Bit.	Bit.	Forestburg(85)	Balmer(82)
Pressure (MPa)	0.1	0.13	4.0	2.5	1.08	0.1	0.1
Temp. (^{o}C)	980	1925	1400	980-1370		903	932
Steam/Oxygen (kg/kg)	0.5	0.4	0.5	4.0	1.0	3.0	3.4
Gas Composition (mole %) CO	48	53	52	18	49.1*	26.3**	29.6**
CO_2	14	10	12	30	17.2	34.5	30.6
H_2	35	36	35	40	29.8	36.0	36.2
CH_4	2	-	-	9	3.2	3.2	3.03
N_2	1	-	-	1	0.3	-	-
Others	-	-		2	-	-	-
Gas Calorific Value (MJ/m^3)	10.7	11.3	11.1	11.3	10.7	8.7	9.1
Cold Gas Efficiency (%)	72	75	75	80		76	65

References: Proustein and Hicks "Synthetic Fuels" McGraw-Hill (1982)

Salvador et al. "The Westinghouse Coal Gasification Process" Int's Gas Res. Conf. June 1980

* Purge free bases ** N_2 - free basis

Table 7 Project Budget and Committed Expenditures 1982-83

	Committed Expenditures (1 April'82-30 April '83)	Contract Budget (1 April '82-31 March '83)
Labour	$	$
1. Principal Investigator	4875	4500
2. Research Associates and Assistants	52,837	59,837
3. Graduate student	3229	9660
4. Workshop Technician	3460	3460
5. Fringe Benefits	3912	3700
Materials, Supplies* Subcontracts	11,340	13,520
Equipment	32,509	10,920
Computer	277	1000
Travel & Living	591	3000
Overhead *	21,727	25,310
	$134,000	$134,905

* Exact figures will be billed by the U.B.C. Finance Dept. Customs and freight items cannot be fixed at this time.

Spouted Bed Gasification of Western Canadian Coals

Work Statement 1982-3

Objectives

1. Extend range of coals gasified to provide broader data bank on behaviour of Western Canadian Coals.

2. Operate spouted bed on oxygen rather than air to produce a synthesis gas instead of a low calorific value gas.

3. Complete the modelling study of the spouted bed gasifier, and with it the experimental measurements of in-bed composition profiles.

4. Study the feasibility of sulphur removal by injection of limestone or dolomite into the gasifier.

5. Provide backup data and information as required for the design study of the 15 tonne/day gasifier.

Major Tasks

1. Four to six gasifier runs will be carried out with a bituminous coal from the Crowsnest coal field. The gasifier is a pilot scale operation and the experimental run frequency is a maximum of three runs in two weeks. The major tasks in carrying out the characterization of one coal are given below, and all the essential tasks must be repeated for characterization of other coals described in subsequent Major Task items. Sufficient coal received in lump form must be crushed and screened to yield about six 45-gallon drums of -6+18 mesh coal feed. This work is carried out by ourselves at the B.C. Research prior to trucking the coal to the gasifier site. Inert bed material must also be crushed and screened. The gasifier is prepared for a run by inspecting thermocouples, installing a new gas filter element, checking the chromatographs calibration, etc., charging the inerts and heating up the reactor with propane. The actual run takes place for a period of 4 to 6 hours, during which gas samples are taken, temperatures and pressures are recorded, and in some cases a series of programmed changes made to the operating conditions. At the end of the run the reactor is allowed to cool overnight. The bed is discharged the following day and the reactor cone removed, the in-bed char screened out, sieved to determine its size distribution, weighed and analysed for combustibles in a high temperature furnace. The coal feed drum is removed and contents weighed. The cyclone receivers are opened up, and the carry-over char weighed, sieved and a combustible determination done. Heat exchangers may be cleaned. Preliminary raw data calculations are done and all the flow rate, temperature, pressure and compositional data taken during the run (typically 12 - 16 determinations of gas composition,

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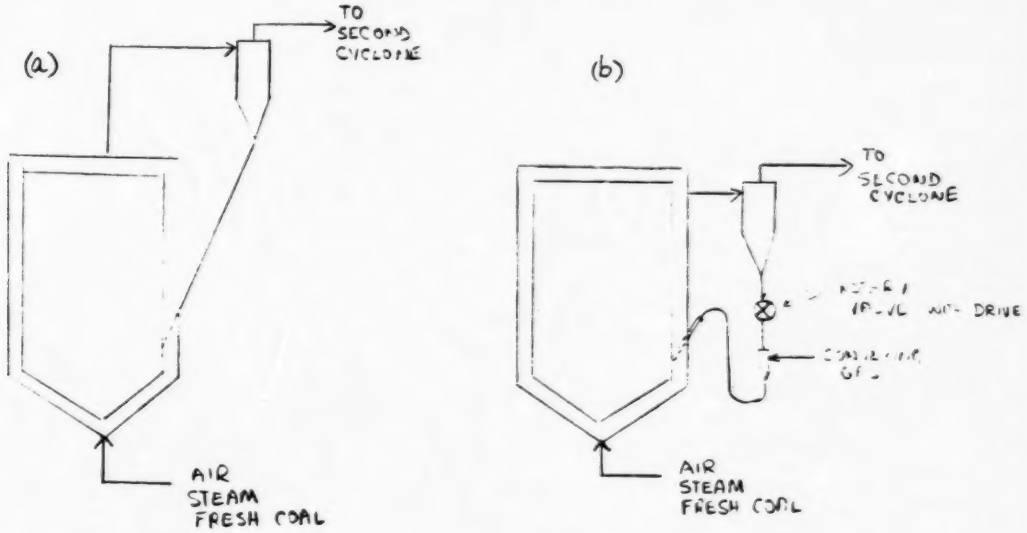
for example,) are punched into the computer for processing. This includes temperature scans in the reactor and gas coolers. Computer output is analysed and plotted, and results of the run assessed. The preparations for the next run are then begun. To characterize a coal, a series of runs would be carried out at different air to coal ratios, and different coal feed rates.

1 April - 30 April

2. Short duration experiments have shown that a gain in efficiency is possible with recycle of product char from the first cyclone to the gasifier inlet. However, even with recycle the carbon content of the waste solids is high. The recycled char appears to blow rapidly through the gasifier if fed in through the high velocity air jet. Feed problems have been experienced. The low head room in the laboratory prevents the use of a direct cyclone/dip leg arrangement as shown in Figure (a). The set-up



shown in (b) will be attempted, in which a star-feeder will be used to meter dust from the cyclone outlet, and conveying gas will be used to introduce the feed into the annulus. This should result in a positive recycle to the gasifier, and the introduction of this feed into the annular region where it can readily react.

To make this modification involves the purchase of the rotary valve, its installation and calibration with char, the welding to the gasifier of a char inlet above the conical section, and the drilling of a suitable hole through the refractory lining, and connection of piping. Once installation and testing are complete a series of three or four runs is planned to determine whether injection of char into the annulus is better than injection into the spout with the incoming air, and to determine the extent of carbon conversion possible with char recycle over

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extended runs of about 12 hours. A good ash balance which has not yet been accomplished with recycle, should be possible.
1 May - 15 June

3. The gasifier which now runs on air and steam producing a low calorific value gas, will be converted to permit operation with oxygen and steam and produce a medium calorific value gas. It is planned to have a dual connection system so that air can be used, for example during heat up, and then the oxygen flow switched on. Thus various combinations of oxygen and air can be used. In order to use essentially the same throughput of coal, and the same reactor lining, the flow of oxygen to the reactor will be the same as with air operation, but steam rather than nitrogen will moderate the temperature so the bed will not reach slagging temperature. The following changes will be made to the unit:

   i) The steam line will be enlarged and a new steam flow meter purchased and installed.
   ii) A fenced-off liquid oxygen tank platform will be constructed outside the Chemical Engineering building, after consultation with the oxygen supplier, and the University Physical Plant and Safety committee. Piping from the liquid oxygen tank will be run up to the laboratory.
   iii) A jet scrubber will be designed, fabricated and then installed between the heat exchangers and the particulate filter. A special water control system will be required to prevent occluded gases from being released into the laboratory water drain.

   Once these changes are made, and the new equipment tested, a series of runs will be made. 16 June - 31 July

4. The spouted bed gasifier will then be evaluated for production of medium calorific value gas. The first tests will be done with a non-caking coal such as Forestburg with which we have considerable experience with air-steam gasification. The effects of steam and oxygen to coal ratios will be of particular interest in these tests. About six runs will be carried out.
1 August - 30 Sept.

5. The system described in Major Task 3 will be used to treat a caking coal, for example, Sukunka in a repeat of Major Task 4. About six runs will be carried out. 1 Oct. - 30 Nov.

6. Tests of oxygen gasification of lignite will be carried out, as described in Major Task 1 and 4. 1 Dec. - 31 Jan.

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7. Sulphur compounds in the coal are released during gasification primarily as H_2S and SO_2. These can be removed by wet-scrubbing process which treat cooled gas, or hot dry scrubbing processes. In another variant either limestone or dolomite can be injected into the bed to adsorb sulphurous compounds. Reactions are not as rapid as under oxidizing conditions however some sulphur removal is possible. In this task, a small feeder will be set up to inject limestone or dolomite particles into the bed to determine the extent of sulphur removal possible in the spouted bed. Local Texada Island limestone and dolomite from the B.C. interior will be tried. Preliminary runs will be restricted to one coal, and tests will probably be attempted first with air gasification. 1 Feb. - 31 March

8. During the course of Major Tasks 1, 2, and 4-7 an on-going program of radial and axial gas composition measurements will be made by graduate student R. Srinivasan. A variety of sampling probe designs are being considered during Fiscal Year 81-82, and some preliminary measurements will be made before F.Y. 1982-3 begins. The data from these measurements will be used as input into a mathematical models of the spouted bed gasifier now being developed. The compositional data will give a clearer picture of what is happening within the reactor, and will provide guidance to selecting the most appropriate model for scale-up calculations. Gases will be analysed for O_2, N_2, CO, CO_2, and CH_4. 1 April - 31 March

9. Current work has shown that a kinetics-free mathematical model using input data of coal feed rate, coal composition, air and steam rates and reactor temperature gives a good prediction of off-gas composition. This model can not predict carbon conversion, temperatures, effect of coal and blast feed rates etc. A kinetic model is being evolved which takes account of the gas flow in the spouted bed and the kinetics of the coal gasification reactions. The reactor is divided into a number of zones and mass and heat balances, fluid flow equations and kinetic expressions are applied. Further development work on the model will be carried out over the year to account for gas mixing and fluid dynamics in the bed and predict the temperature field in the reactor, the carbon conversion and the product distribution. 1 April - 31 March

10. Write a Final Report 1 March - 31 March

Proposal: Gasification of Western Canadian Coals in a Spouted Bed-Program 1983-4

Objectives

1. The simplicity of operating the spouted bed gasifier with a char-ash bed rather than with inert particles would be a strong advantage. Although preliminary tests have shown that the gasifier can be run in beds of char, in 1983-4 a series of longer-term runs with recycle in this mode will be attempted to demonstrate feasibility with caking and non-caking coals.

2. Operating with char rather than inerts provides some flexibility for ash removal. Tests are planned to determine the feasibility of operating in the ash-agglomerating mode rather than the dry ash mode.

3. To date most of our work has been done on closely-sized coal size fractions. In a commercial application it is not feasible to remove fines from the feed without economic penalty. In 1983-84 most of our work will be done with wider sized coal fractions e.g. - 4.76 mm, and no fines removal to assess fully the effect of fine particles in the feed.

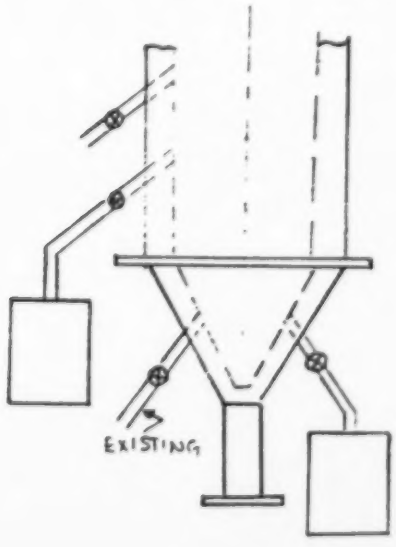
4. Work will be continued on the deleterious raw gas constituents (particulates and sulphur species), to build up a solid bank of data for design of removal processes.

5. Preliminary experiments will be run to test the feasibility of applying catalysts to gasify the less-reactive bituminous coals.

6. Work will be continued at the graduate student level on measuring composition and temperature profiles within the spouted bed gasifier. This work supports our mathematical modeling for scale-up.

Major Tasks

Task 1. In order to prepare for operation with the char bed, and with high ash coals, or for particularly long runs, provision for solids overflow will be added to the gasifier. One or two holes will be drilled through the shell and refractory, and an insert made to remove solids by gravity flow. A sealed receiver will be provided.



The basic design will be based on our bed sampling system in the cone. This will permit firm control over bed level in dry ash operation. During this modification, part of the top section of the gasifier will be re-lined, and provision made for removal of agglomerated ash from the cone (See Task 4).

1 April - 30 April

Task 2. The first experimental task is to assess the performance of the spouted bed gasifier operated with char rather than inerts using a non-caking sub-bituminous coal (Forestburg). This would involve some 4 or 5 runs of 6-12 hours duration. Each run would consume about 4 drums of coal and entail 3 days crushing and screening.

The reactor would be operated as follows: About 13 kg of fresh coal would be charged to the gasifier, and heated under spouting conditions to 550°C with combustion products from the propane burner. The propane flow is reduced, and air is increased as coal feeding is commenced. When the operating temperature of about 850°C is attained, and conditions become steady experimental data such as temperatures, pressures, solid and gas flowrates, are recorded and gas samples are taken at regular intervals of 15-minutes over 4-10 hours. At the end of the run air, steam and coal are shut off, nitrogen is used to purge the system and their reactor is allowed to cool overnight. The following day the cyclone dust receivers are weighed and representative samples analyzed. The dust filter is removed from its drum, weighed and the solids analyzed. The reactor cone is lowered, and the char-bed removed, weighed and analysed for particle size distribution and loss on ignition. Data collected during the run are assembled and punched into the computer. Results are analyzed and plotted.

Additional runs to confirm satisfactory operation with the char bed will be attempted with oxygen rather than air gasification, and at various throughputs. Experiments would be run in the dry ash mode, i.e. with excess steam when oxygen is used. Running conditions would be selected to match to the extent possible available runs with the inert silica rather than the char bed.

The first two runs will be carried out with sized −4 +16 mesh coal. Further work will be done using a wider cut of feed coal with no fines removal e.g. −4 mesh.

1 May - 15 June

Task 3. Task 2 will be repeated with a bituminous coal Balmer or Sukunka coal will be used. Again, 4-5 runs are planned with variations in coal size distribution, throughput and with oxygen operation.

15 June - 30 August

Task 4. In experiments up to this point the gasifier has been run in the dry ash mode. Ash leaves the gasifier conveyed in the gas. An attempt will be made to operate the gasifier in the agglomerating ash mode again using a char bed. This will require higher temperatures in the cone section, which will be achieved with some oxygen enrichment of the air and at relatively low steam rates. Some trial and error is anticipated to obtain satisfactory agglomeration, and removal of the ash. Preliminary experiments may be done in batch rather than continuous operation.

Modifications may be required to withdraw agglomerated ash near the cone bottom, however initial attempts will be made through the existing bed sampler coupled with a new hot ash receiver.

1 September - 15 October

Task 5. Assuming Task 4 is successful, a series of four or five comparative runs in the ash agglomerating mode will be run over a range of conditions with both caking and non-caking coals.

15 October - 15 January

Task 6. A preliminary series of tests of catalytic gasification will be carried out. In three or four experiments coal will be either pretreated with different levels of K_2CO_3 solution and the coal permitted to dry at near room temperature, or the solution will be sprayed directly into the cone section. Short gasifier runs will be made to determine potential increase in gasification efficiency, gas heating valve or carbon conversion with the catalyst. These experiments will be done on less-reactive bituminous coals which are relatively difficult to gasify.

15 January - 15 March

Task 7. Further efforts will be made to quantify deleterious constituents in the gas. This task which will cover work over several months will involve an examination of cyclone efficiency, characterization of particulates from the downstream cyclone receiver, and from the dust filter. If desireable a hot gas clearing device could be tested. Samples of dusts will be sent to the Energy Research Labs for size distribution by coulter counter. Sulphurous species in the gas will be monitored.

1 April - 1 February

Task 8. Internal bed gas composition and temperature data will continue to be gathered by graduate student Z. Haji-Sulaiman. This data will be incorporated into mathematical models of the spouted bed gasifier.

1 April - 1 December

Task 9. Assemble data and write up report.

1 - 31 March

Budget Estimated for 1983-84

A. Labour at firm rates and payroll cost.

1. Principal Investigator
   Dr. A.P. Watkinson — Est. $4,770.00
   Firm $198.75/day for est. 24 days

2. Senior Research Engineer
   G. Cheng — Est.$29,256.00
   est. $2438/month est.
   12 months

3. Junior Research Engineer
   S. Low — Est.$23,532.00
   est. $1961.00/month for
   est. 12 months

4. Graduate Student — Est. $9,660.00
   est. $805/month for est
   12 months

5. Workshop Technician — Est. $3,700.00
   est. $16.35/month for est
   211.5 hr (Construction of bed overflows, ash agglomeration systems)

6. Fringe Benefits on A1 to A5 — Est. $3,545.90
   @ est. 5%
   Subtotal $74,463.90

B. Materials and Supplies
   (Gases including oxygen, thermocouples, chromatography supplies, upper section refractory re-line, bed sampling probles) — $6,000.00

C. Equipment
   Bed overflow system; ash agglomerating system — $3,000.00
   Data-logger (to be shared with other projects) — $5,000.00

D. Computer at University Charge Rates — $1,000.00

E. Subcontracts
   1. Coal, freight, coal preparation, coal analysis — $7,000.00
   2. Oxygen tank system rental — $1,500.00

F. Travel and Living
   Travel to coal conference — $3,000.00
   (G. Cheng to Coal Technology '83, P. Watkinson to Energy Technology '84 or other)

G. Overheads to UBC
   30% of Items A, and B, 2% of F — $24,200.00
   24,139.17 + 60

$125,163.00

APPENDIX B

Experimental Gasification Data 1982-83

TABLE B-1 Experimental Gasification Runs 1982-83

Run No.	Coal	Coal % H_2O	Coal d_p (mm)	Gasn Mode	FEEDRATE* (kg/h) m_{O_2}	m_C	m_A	m_S	m_{CH}	m_{ox}/m_C	m_S/m_C	TEMP °C AVG	MAX	Gas Heat. Val. MJ/m³	Gas Yield m³/kg	Gas Analysis (%mol) H_2	CO	CO_2	CH_4	N_2	% C Conversion	Cyclone Catch kg/h	Cyclone Catch % C	η_{HOT} (%)	% H_2O Decomp.	Run Duration hr:min
62a	Balmer Reject	1.90	2.00	Sp		18.52	99.45	6.73	—	1.25	0.36	904	946	1.38	4.45	4.59	4.62	13.67	0.71	76.46	51.89	11.36	21.58	40.89	18.15	3:23
62b	Balmer Reject	1.90	2.00	Sp		29.43	102.92	10.75	—	0.81	0.36	863	874	2.22	3.03	7.02	6.11	12.92	1.46	72.39	38.52	11.36	21.58	36.20	19.08	3:28
62c	Balmer Reject	1.90	2.00	Sp		36.09	109.41	10.51	—	0.68	0.29	868	881	2.13	2.99	7.46	6.08	11.55	1.30	73.61	29.00	11.36	21.58	29.19	18.30	3:23
63a	Balmer Reject	1.90	2.00	Sp		27.51	101.56	10.09	—	0.86	0.37	907	936	1.82	3.33	6.00	5.98	14.68	0.98	72.36	43.09	33.85	30.90	35.99	17.72	2:23
63b	Balmer Reject	1.90	2.00	Sp		29.93	103.94	10.61	—	0.82	0.36	879	908	1.99	3.20	7.39	5.85	15.99	1.04	69.80	44.97	33.85	30.90	36.18	21.20	2:23
64a	Balmer Reject	1.90	2.00	Sp		32.38	90.41	4.79	—	0.65	0.15	820	872	1.79	2.98	6.98	5.83	15.95	0.63	71.11	33.38	38.89	35.73	26.56	17.32	2:59
64b	Balmer Reject	1.90	2.00	Sp		26.83	98.51	9.17	—	0.87	0.34	899	997	1.75	3.34	6.26	6.89	15.86	0.45	70.59	47.58	38.89	35.73	37.09	16.63	2:59
64c	Balmer Reject	1.90	2.00	Sp		29.71	100.46	9.91	—	0.95	0.40	679	910	1.76	3.68	7.26	7.10	15.73	0.45	69.83	51.87	38.89	35.73	35.17	20.09	2:59
65a	Balmer Reject	1.60	2.00	F		17.95	95.19	3.66	—	1.23	0.20	756	801	1.20	9.30	2.62	9.67	13.28	0.89	73.58	49.69	39.85	39.76	39.22	10.72	2:16
65b	Balmer Reject	1.60	2.00	F		26.56	96.15	5.97	—	0.89	0.20	709	769	1.06	2.93	2.65	3.09	19.59	0.89	72.52	32.71	39.85	39.76	22.92	7.22	2:16
65c	Balmer Reject	1.60	2.00	F		28.21	93.53	5.89	—	0.80	0.20	758	870	1.55	2.90	4.19	4.77	13.99	1.10	75.63	36.42	39.85	39.76	27.07	11.71	2:16
66	Forestburg	23.50	2.06	Sp		34.08	100.00	12.87	—	0.68	0.55	809	822	3.51	3.60	13.84	8.11	25.76	1.13	51.55	71.72	5.87	11.59	55.05	37.89	2:06
67a	Rosetyn	27.01	1.83	F		27.65	105.62	16.04	—	0.89	0.58	731	737	2.80	3.77	10.98	7.29	15.92	1.56	63.58	64.13	18.76	39.31	39.88	31.32	2:30
67b	Rosetyn	27.01	1.83	F		31.92	101.89	18.51	—	0.79	0.58	709	716	3.84	3.68	13.62	6.15	17.94	1.70	62.27	60.09	18.76	39.31	39.33	36.54	2:30
68	Rosetyn	27.01	1.83	F		25.85	108.15	16.40	—	0.97	0.63	730	816	3.10	4.38	15.29	9.98	15.19	0.18	59.69	75.53	7.70	44.23	43.01	48.86	2:25
69	Balmer Reject	1.60	2.00	F		26.96	99.15	7.28	—	0.86	0.27	700	706	1.86	3.32	6.53	5.31	17.15	1.12	69.69	48.39	32.89	44.35	33.22	18.92	1:25
70	Balmer Reject	1.60	2.00	F		24.99	87.96	1.00	—	0.83	0.04	663	930	2.72	3.56	7.60	9.69	18.03	1.70	63.04	64.25	25.48	47.60	38.16	31.78	1:12
71a	Forestburg	23.50	2.06	Sp		32.28	91.66	10.96	0.81	0.71	0.34	782	786	3.58	3.69	13.93	12.33	15.86	1.53	56.35	68.04	6.82	31.58	57.21	45.00	3.56

* m_C COAL, MAF BASIS ; η_{HOT} = HOT GAS EFFICIENCY ; SP = SPOUTED BED F = FLUIDIZED BED , $m_{ox} = m_{O_2} + 0.23\, m_A$ m_{O_2} = PURE OXYGEN

TABLE B41 Experimental Gasification Runs 1982-83 (continued)

Run No.	Coal	Coal % H_2O	Coal d_p (mm)	Gasn Mode	FEEDRATE* (kg/h) m_{ox}	m_c	m_A	m_s	m_{ch}	m_{ox}/m_c	m_s/m_c	TEMP. °C AVG	MAX	Gas Heat. Val. MJ/m³	Gas Yield m³/kg	GAS ANALYSIS (%mol) H_2	CO	CO_2	CH_4	N_2	% C Conversion	CYCLONE CATCH kg/h	% C	η_{HOT} (%)	% H_2O Decomp.	Run Duration hr: min
71b	Fordbury	23.50	2.06	Sp		32.23	98.17	10.96	2.08	0.71	0.16	789	792	3.48	4.00	14.42	14.06	18.09	1.96	55.18	71.65	6.82	31.52	62.88	44.78	3:56
72	Fordbury	23.71	2.06	Sp		31.07	94.94	11.76	3.35	0.75	0.38	759	771	3.20	3.40	14.02	11.83	15.25	1.97	57.02	68.29	12.11	50.05	61.18	44.57	6:00
73	Fordbury	23.71	2.06	Sp		27.05	96.84	12.71	—	0.83	0.47	792	821	3.81	3.09	13.89	16.20	17.14	0.87	58.30	72.59	7.81	28.19	58.60	41.10	3:55
74	Fordbury	23.71	2.06	Sp			Unsuccessful				Run															0:22
75	Bulmer	1.97	2.00	Sp		31.58	95.33	6.69	—	0.59	0.18	821	841	3.00	2.83	9.89	10.90	12.76	1.45	65.50	34.19	24.01	58.51	29.90	29.70	2:56
76b	Bulmer	1.97	2.00	Sp	15.50	39.97	28.20	15.26	—	0.63	1.15	801	816	7.13	1.97	19.37	24.52	17.52	6.99	32.95	95.00	22.12	68.39	48.80	8.97	3:56
76b	Bulmer	1.97	2.00	Sp	15.90	43.06	27.00	12.13	—	0.52	1.69	801	809	7.86	1.91	21.87	22.16	19.36	6.69	29.12	92.66	22.12	68.39	59.40	9.96	3:56
77a	Fordbury	23.71	2.06	Sp	16.70	36.99	27.00	9.18	—	0.63	2.52	761	781	5.50	1.72	30.68	9.17	31.32	1.67	27.20	98.32	11.29	41.99	52.90	13.39	3:97
77b	Fordbury	23.71	2.06	Sp	17.60	43.95	27.50	88.97	—	0.56	2.07	710	716	5.40	1.62	29.17	8.91	33.59	2.03	26.25	91.89	11.29	41.99	45.70	14.81	3:97
78	Bulmer	1.97	2.00	Sp	15.70	37.06	28.80	11.35	4.34	0.60	2.09	780	829	6.12	1.33	15.72	23.17	20.08	4.96	36.51	35.26	22.27	61.81	36.60	5.66	2:49
79	Bulmer	1.97	2.00	Sp	19.30	35.33	40.80	12.62	5.58	0.82	2.06	824	856	6.27	1.95	18.96	19.95	20.95	5.81	34.33	51.00	17.92	18.26	50.10	9.52	3:35
80	Bulmer	1.97	2.00	Sp	19.10	39.91	39.90	12.57	5.58	0.83	2.08	863	868	5.36	1.88	21.72	17.08	29.91	2.21	29.58	56.98	11.91	60.96	41.90	14.27	2:59
81	Bulmer	1.97	2.00	Sp		31.33	111.96	7.84	4.09	0.70	0.31	881	912	3.08	2.82	9.69	10.61	16.18	1.63	61.23	37.02	21.19	57.31	35.69	32.40	3:35
82	Bulmer	1.97	2.00	Sp	31.80	31.53	13.00	96.95	5.91	0.99	2.51	853	815	8.50	1.72	29.89	28.32	31.16	3.56	7.07	61.93	15.86	65.16	58.75	13.10	2:50
83	Bulmer	1.97	2.00	Sp	28.70	38.89	13.50	109.50	5.32	0.82	2.79	917	932	8.27	1.68	32.88	26.90	27.80	2.75	9.10	63.09	14.98	69.19	65.12	15.28	3:32
84a	Bulmer	1.97	2.00	Sp		31.32	91.38	7.96	5.18	0.61	0.30	836	845	5.01	2.50	9.96	9.05	12.87	1.81	66.35	39.63	22.08	67.97	31.00	30.84	3:41
84b	Bulmer	1.97	2.00	Sp		31.32	99.69	7.96	—	0.62	0.30	890	895	3.22	2.37	12.99	10.39	13.55	1.71	61.95	38.70	22.08	67.97	35.53	41.09	3:41
85	Fordbury	23.71	2.06	Sp	26.80	31.93	12.90	96.68	1.92	0.93	3.03	881	903	7.50	2.35	30.98	22.65	29.71	2.68	13.98	86.59	9.30	18.68	83.98	16.35	1:46

* m_c COAL, MAF BASIS ; η_{HOT} = HOT GAS EFFICIENCY ; SP = SPOUTED BED F = FLUIDIZED BED $m_{ox} = m_{O_2} + .25\, m_A$ m_{O_2} = PURE OXYGEN